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ABSTRACT

To better understand the multiphase hydrodynamics in flotation deinking processes, we have used gamma densitometry to measure local and global gas holdup in bubble columns with aqueous suspensions of recycled newsprint. Flow systems included air rising through stagnant suspensions and cocurrent flow of air and pulp. We examined suspensions of several consistencies, including pure water, with a variety of injected air flow rates. The effect of surfactant was also examined. The results are applied to interpret known characteristics of typical flotation cells.

We find that fibers promote the coalescence of bubbles which can significantly alter the bubble size distribution. The presence of a fiber network retards the rise of bubbles, which by itself would decrease the bubble rise velocity and increase holdup. On the other hand, the retarding of small bubbles promotes coalescence into larger bubbles that can rise more rapidly through the network. The net effect can be a decrease or increase in gas holdup. For most of the conditions we observed in stagnant pulp suspension flows (gas flow only), fibers promoted a decrease in holdup

compared to pure water. In cocurrent flows of gas and pulp suspensions, higher gas holdup may be achieved in pulp suspensions than in pure water, though further tests are needed to confirm this result.

In stagnant pulp suspensions, the fiber network also promotes channeling of the air, which results in loss of interfacial and poor ink removal. The transition to channel flow occurs at a critical gas velocity which decreases with increasing pulp consistency. For consistencies of 2% and above, channeling occurs even at very low gas velocities.

Recirculatory flow in a column or tank is easily established in the buoyant flow system of flotation deinking. The tendency is for the gas to rise toward the center of a column (or local buoyant cell) with liquid flow returning downward at the walls of the column (or at the edges of buoyant cells). This recirculation may be undesirable, for it can entrain ink-laden foam back into the suspension and can also backmix the deinked pulp into the dirty pulp. Ideas to prevent backmixing and to prevent channel flow (loss of interfacial area) have now been proposed and will be tested further in the future.

BACKGROUND

Flotation Deinking

A critical challenge facing the industry in expanding the use of recovered paper is the need to efficiently remove ink from post-consumer waste. Flotation deinking, in particular, will be an increasingly common operation in recycling operations. Unfortunately, most of the development of flotation deinking for the pulp and paper industry has relied on trial-and-error development, and much of the design work for flotation hydrodynamics (including claims relating to bubble size) has been based on experiments in clear water flows. The effect of a fibrous slurry on the bubble dynamics or fundamental is unknown. Even the most rudimentary issues in three-phase fiber suspension flows seem to have been neglected in the literature, resulting in a serious lack of knowledge about the behavior of a critical operation.

In flotation deinking, bubbles are formed in a fiber suspension which can attract hydrophobic ink particles. The bubbles coalesce and rise, carrying the ink to the surface where steady removal occurs by skimming or overflow. Chemicals are added to help attach the ink particles on the bubbles. Although at least six different vendors produce a variety of flotation deinking systems, all have several basic features in common. A flotation deinking cell begins with a mixing zone, where a fast-moving fiber suspension stream is mixed with air. High interfacial area is desired at this point to collect ink particles. The air volume fraction (at atmospheric pressure) is typically less than 30%. Residence time in the mixing zone is on the order of a few milliseconds, while total residence time in the flotation cell may be one minute or more. Following mixing, the three-phase mixture enters the coalescence zone, where much less shear is present. The bubbles interact and coalesce, forming bubbles on the order of millimeters or larger which have sufficient buoyancy to rise through the fibrous network. Finally, there is the separation zone, where large bubbles (ca. 1 cm or greater) are present, and where the ink-laden froth at the surface is removed. As few fibers as possible should be removed.

Some flotation cells have all three zones in one large tank. Others attempt to separate mixing from coalescence and separation. Some cells have strong swirling flows induced by tangential injection of pulp or gas, while others have primarily vertical or horizontal suspension flows. A few employ moving parts to induce strong mechanical shear to create fine bubbles.

Much of the engineering of flotation deinking systems has been based upon flotation processes in the mining and minerals industries, where flotation has been used for many years to separate minerals based on surface properties. However, systems for deinking have evolved to have significantly different flows, making mineral flotation concepts inapplicable in many cases. Mineral flotation cells operate at low Reynolds number (on the order of 2000 or less) in the mixing zone, while the Reynolds number in a deinking cell (based on the dimensions of the mixing zone) can be as high as 10^6. Fiber suspensions are also much different in their behavior than the relatively simple mineral suspensions subject to flotation. Fiber-fiber entanglement and flocculation, bubble-network trapping, and fiber drag reduction are examples of the effects that may be important in flotation deinking and that are alien to mineral flotation.

Based on extensive mill experience, pilot trials, and some academic efforts, the macroscopic behavior of some flotation cells is generally understood at an empirical level. For example, it is known that ink particles in the range of 20-100 μm can be removed with high efficiency. It is also known that different pulps have different critical concentrations at which deinking efficiency plummets. The effects of chemicals, ink type, air injection rates, temperature, and other properties have been examined in terms of collection efficiency or brightness of the resulting pulp. However, based on our discussions with manufacturers and users of flotation cells, as well as several literature searches, it is clear that little is known about the behavior of gas bubbles in a fiber suspension.

Vendor claims concerning air bubble size distribution and bubble dynamics in these systems are based on measurements in easily studied gas-water systems; fundamental information for bubbles in pulp suspensions does not appear to be available. Of particular concern is the effect of fibers on bubble size distribution, the transient aggregation of bubbles in the system, and nonideal flow patterns such as channel flow.

Features of Multiphase Flow

Due to its numerous applications, gas-liquid two-phase flows have been intensely studied over several decades. Three-phase flows involving liquid, solid, and gas have received increasing attention recently, but have been studied far less than two-phase systems. Key topics of investigation in gas-liquid flows have been two-phase flow patterns, bubble size, pressure drop, and heat transfer between the two-phase mixture and solid surfaces (e.g., 1,2). The field, however, is essentially empirical, and research efforts aimed at developing mechanistic models are underway worldwide.

The flow pattern, or phase topology, is an issue of great importance. When gas and liquid are both present in a flow system, a variety of flow patterns are possible (3). For an example of relevance to the present study, we will consider the case of a vertical column of liquid with upward flow of both liquid and gas. If the superficial gas velocity (volumetric flow rate divided by cross-sectional area) is low and the liquid velocity is also low, and if the gas is introduced as finely dispersed bubbles, a steady bubbly flow regime may exist with individual bubbles rising in the liquid, having relatively minor interactions between the bubbles. In the bubbly regime, the liquid is the continuous phase and the gas is discontinuous (the interconnected of phases and the geometry of the interfacial area - the phase topology - are crucial concerns for mass and heat transfer processes). Increasing the gas flow rate may result in larger bubbles that aggregate and break up and create obvious turbulence in the liquid. This may be called bubbly-turbulent flow or churn-turbulent flow. A further increase in gas velocity may result in slug flow, with large gas bubbles that nearly fill the width of the column. For wide columns, however, larger bubbles and more churning fluid motion is observed instead of slugs. When a slug flow regime exists, a further increase in gas flow rate may result in annular flow, with the gas establishing a continuous "chimney" in the center of the column and the liquid present as a film along the wall. Some authors report a transition regime between slug flow and annular flow featuring liquid waves that break away from the wall and lead to chaotic mixing. This regime has also been called churn flow, or following the terminology of Hewitt and Jayanti (4), churn flow of the third kind, though it may truly be a form of slug flow and not a separate pattern. At very high gas velocities, the liquid may be dispersed into fine drops or wisps in the gas phase. The patterns mentioned here will be functions not only of flow rates of both gas and liquid, but of initial flow conditions, downstream flow conditions, flow geometry, fluid properties such as surface tension, density, and viscosity, etc. For gas-liquid flow in vertical tubes, Taitel et al. (5) discuss the mechanisms leading to transition from one regime to another and present several flow maps. For an extensive set of flow regime maps for air-water flows, see Spedding and Nguyen (6).

Figure 1 shows that the relationship between gas holdup and air flow is a strong function of how the bubbles are introduced. This figure is based on a graph from Zuber and Hench (7) who injected air through perforated plates into a column of water. (For clarity, lines have been drawn instead of individual data points – the actual data show significant scatter in the transition zone for the upper curve.) A variety of orifice sizes and spacings were used. With a few large orifices, the bubbles induced churn-turbulent flow even at low gas flow rates. With many fine orifices, nearly ideal bubbly flow could be achieved, with a transition to churn-turbulent flow occurring with a gas superficial velocity beginning 5 cm/s. (A common but rough general rule for vertical columns with quiescent water is to assume bubbly flow for gas fluxes below 10 cm/s and churn-turbulent flow for higher flows.)

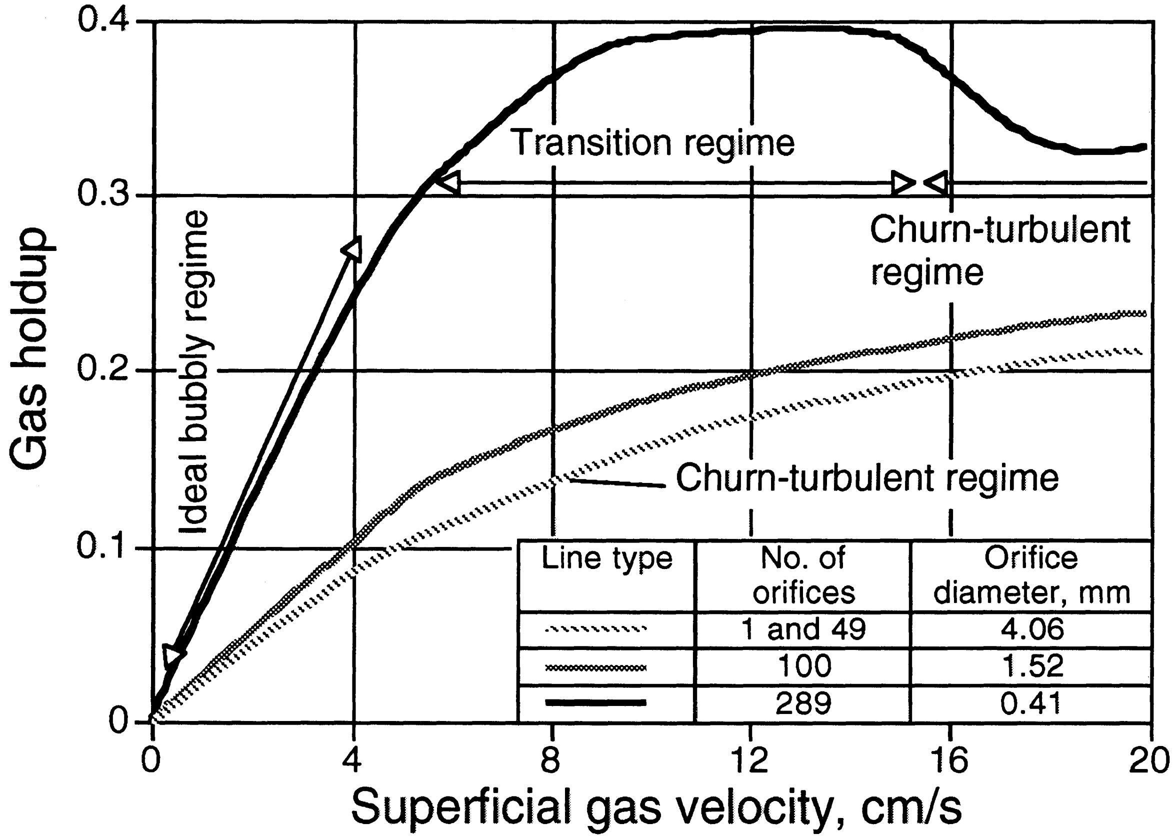

Figure 1. Effect of initial conditions on gas holdup for perforated plate injection of air bubbles into a water column (7).

The behavior of a three-phase flow is significantly different from gas-liquid and solid-liquid two-phase systems, and methods associated with two-phase flow generally cannot be applied to the analysis of a three-phase system. In a three-phase bubble column, for example, the suspended solid particles influence the gas and liquid holdups (volume fraction occupied by a phase) in a complicated manner. At low gas velocities, the gas holdup may decrease with increasing solid concentration, particle size, and solid-liquid density difference (8). Small particles in particular are likely to promote bubble coalescence and decrease the gas holdup, while large particles may promote bubble break up and slightly increase holdup. The influence of solid particles on gas holdup is reduced at high gas velocities.

A variety of techniques have been developed for two-phase and three-phase flow studies, some of which may have potential in fibrous suspensions. Local phase characterization of two-phase flow (transient measurement of the presence of gas or liquid at a point) has been done primarily with: (1) electrical probes, based on measurements of resistance or impedance, to detect the presence of gas in a suitable liquid; (2) optical probes, including fiber optic probes with a wide variety of terminating geometries, to exploit refractive index differences between gas and liquid; and (3) thermal probes, which differentiate between gas and liquid by differences in thermal conductivity. A thorough review of these approaches is given by Saxena et al. (9), who discuss the usefulness of these techniques for coal slurry bubble reactors.

Chemical tracers, especially using the gas phase as a tracer, have some potential in examining gas distribution in a flow system if adequate sampling methods are devised. Flash x-ray radiography offers a two-dimensional stop-motion visualization of density gradients in the flow (10), while gamma-densitometry provides useful time averages of density or void volume along a line through the flow.

A good discussion of three-phase hydrodynamics in vertical flow systems (bubble columns) is presented by Kara et al. (11), who examine mixtures of air, water, coal and mineral ash in a 15.2-cm bubble column as functions of particle size, solids concentration, and liquid and gas velocity. An excellent example of three-phase flow characterization for oil, water, and air systems was recently given by Açikgöz et al. (12).

Three-phase Pulp Suspension Flows

Three-phase flows have been studied by a variety of authors, but three-phase pulp suspension flows are poorly understood. The most relevant research we have found is that of

Walmsley (13), who used a stagnant, vertical bubble column to examine a few cases of bubble rise in dilute pulp suspensions. Pulp was suspended in water and also in clove oil, which has a refractive index near that of cellulose, allowing improved visualization of rising bubbles. Gas holdup as a function of gas flow rate was measured, and the characteristics of bubble motion were observed visually. Walmsley observed bubble flow, channel flow and churn turbulent flow. Network strength was proposed as an important factor governing the interaction of bubbles and flocs. He found that fibers can promote bubble coalescence and can lead to decreased gas holdup relative to pure water, while very dilute (ca. 0.05%) fiber concentrations may lead to an increase in holdup. Long fibers were more effective in promoting bubble coalescence than short fibers.

Robert Pelton at McMaster University in Ontario has studied the effect of flocs on restricting the upward motion of bubbles (14). He has determined the critical bubble size required to break through fibrous networks of varying fiber types and consistencies. His efforts provide some insight into how fibers can cause bubbles to coalesce.

Pan et al. (15) have sought to model the interactions of air bubbles in dilute flotation deinking systems. Stokesian dynamics were applied to treat bubble-particle collisions. More extensive information about macroscopic hydrodynamics will be needed to make this approach more applicable to flotation deinking.

EXPERIMENTAL APPROACH

Gamma densitometry is the basis of our experimental methods. We use an Americium 241 gamma source and an EGG Ortec counter system to measure the attenuation of a gamma beam caused by the presence of mass between the beam and the detector. The logarithm of the gamma counts obtained per unit time is linearly related to the mass present between the source and the detector, with more mass resulting in fewer counts. Using a beam approximately 3 mm in diameter, we can scan across multiphase systems to determine the local gas content at various positions. The measured void volume is averaged along the beam path and is averaged over the time of the measurement, which may range from 10 to 60 seconds. Calibration is required to take into account background radiation levels and variation in wall thickness or other systematic sources of error.

The number of measured counts per minute minus the background counts per minute will be designated as N. The count rate through the centerline of the water-filled column is N_{full}, and the count rate through the empty column is N_o (all measured rates have the background count rate subtracted). In a column with pulp and air present, the void volume (gas holdup) through the center of the column at any vertical position is related to the observed count rate, N, by

$$\varepsilon = 1 - \frac{\ln(N/N_o)}{\ln(N_{full}/N_o)}$$

The present study employed two flow systems for multiphase studies. One system has cocurrent flow of both pulp and air, while the other is a bubble column that has air flow only through a stagnant pulp suspension. The bubble column (stagnant pulp) is a 10-cm i.d. plastic column with three sintered brass air injectors located at 120° increments around the base of the column, as shown in Figure 2. The column is filled with 3 liters of pulp suspension and is placed in the path of a gamma ray beam. At a given air flow rate, as measured by a Hastings mass flowmeter, the column is moved vertically through the gamma ray beam to obtain centerline gamma scans through the cylinder. This provides us with gas content as a function of column height. We can also leave the system fixed at a particular location as we vary the air flow rate.

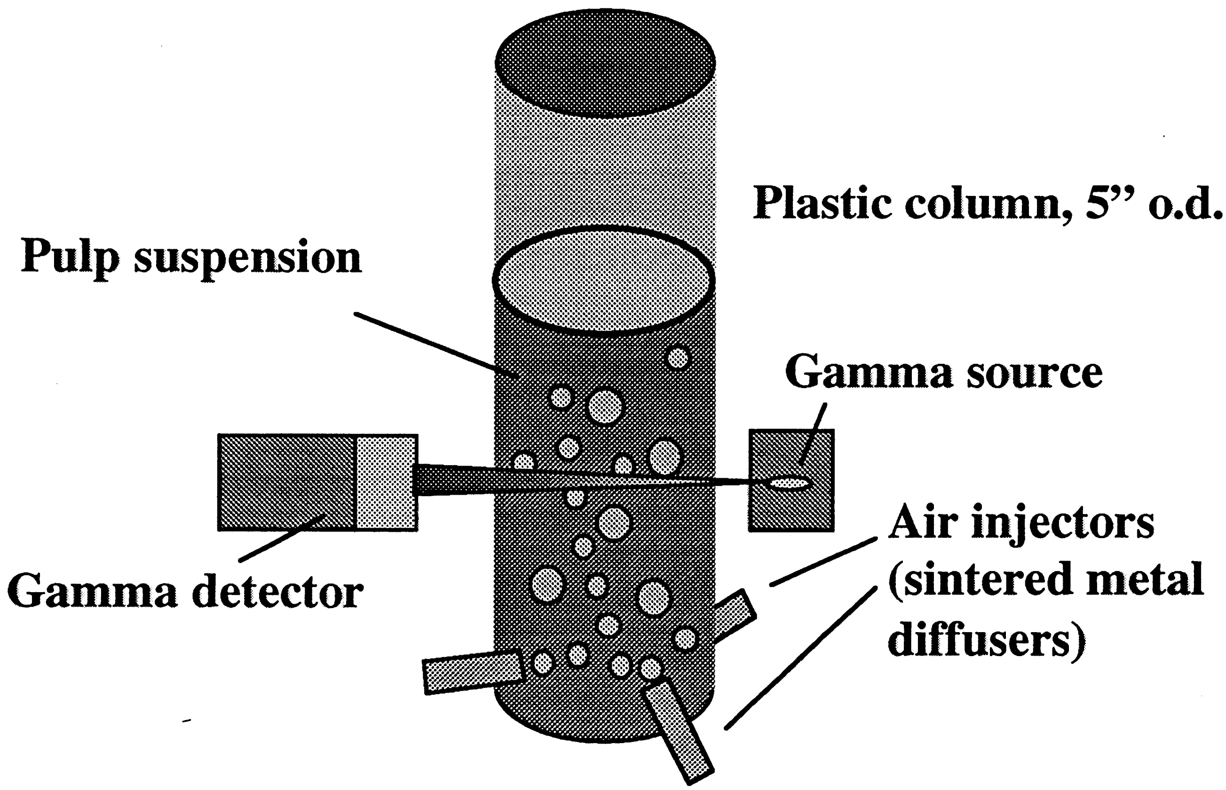

Figure 2. Bubble column with sintered-metal air injectors.

The cocurrent flow device, shown in Figure 3, uses shear in the flowing pulp to disperse injected gas stream. Mixing occurs in a 2.5-cm tube with Reynolds numbers on the order of 10^4 or greater just prior to a conical expansion to a 13-cm clear plastic tube. The tube is equipped with a device that allows the gamma system to scan in both radial and vertical directions, permitting two-dimensional plotting of chord-averaged void volume in pulp suspensions. The gamma system can scan radially and vertically over the pipe. Pulp flow rates are controlled with a Diskflow™ pump and monitored with a magnetic flowmeter. Gas flow rates are monitored with a Hastings mass flowmeter. Air-containing pulp is returned to a tank where most of the air leaves the pulp before the flow is returned to the pump. Superficial liquid velocities range from 0 to 10 cm/s, and gas velocities typically range from 0 to 3 cm/s (higher gas velocities pose problems for steady operation due to excessive air entrainment in the holding tank).

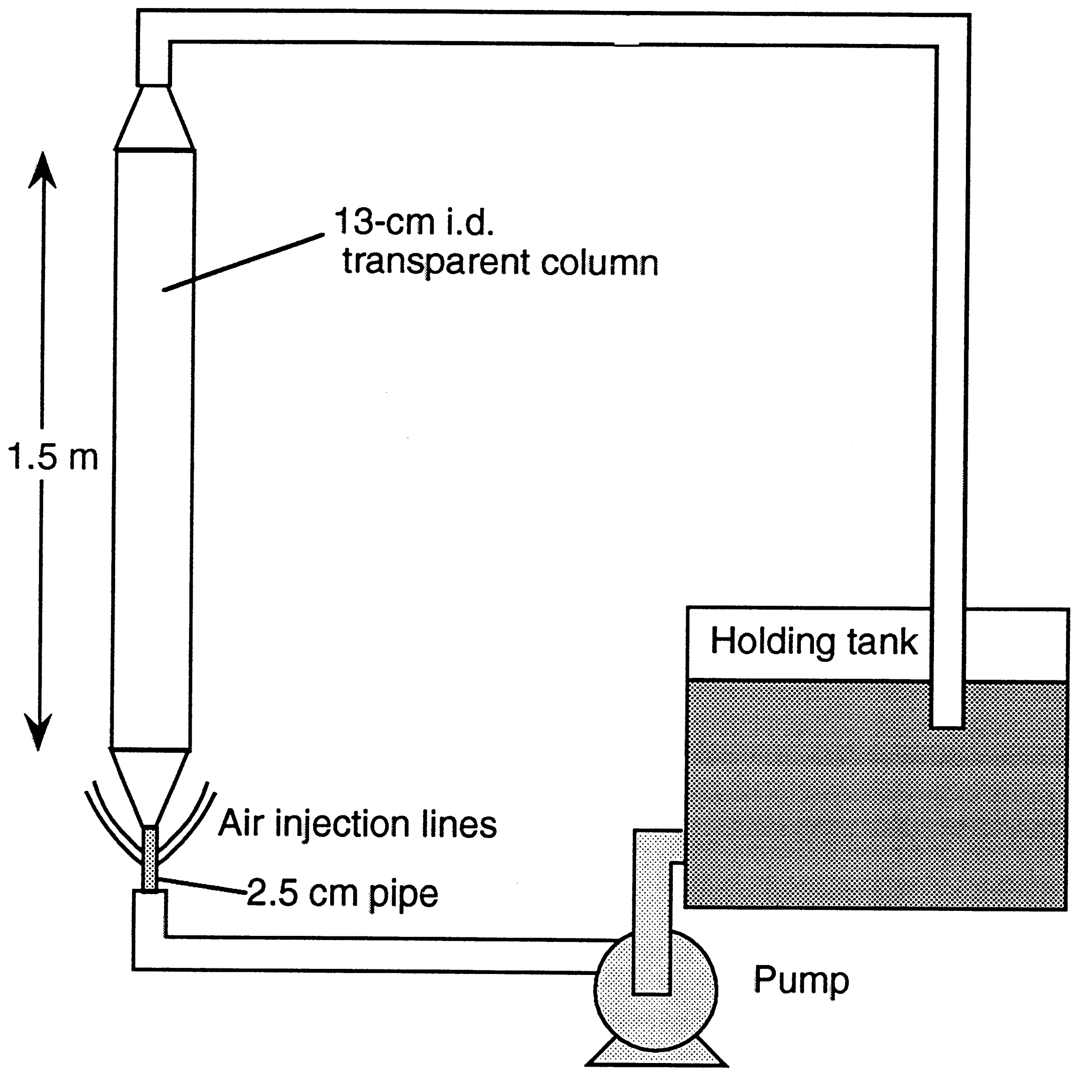

Figure 3. Cocurrent flow device for multiphase flow studies.

While both columns are more narrow than typical flotation cells, wall effects are not expected to predominate the hydrodynamic results. A study of Jones and Ellis (16) in gas-liquid bubble columns indicates that gas holdup is not a function of column diameter for columns greater than 7.5 cm in diameter. Gas holdup is increased by wall effects for more narrow columns.

In all work to date, we have focused on repulped southern newsprint as the pulp suspension. Triton X-100 has been used in studies of surfactant effects.

RESULTS

Bubble Column with Quiescent Pulp Suspension

Even though the fiber network system presents a barrier that slows the rise of an individual bubble, by the same token it holds bubbles in place until enough have coalesced to break through the network. Coalescence can lead to large bubbles that rise rapidly through the stock, rapidly enough to decrease the overall gas content compared to pure water or more dilute stock. One example of this is shown in Figure 4, where we present diameter-averaged holdup measurements (the gamma scan was done through the centerline of the column, along a diameter) for various pulp consistencies, all taken at a vertical position of 22 or 24 cm above the gas injection plane. The results in Figure 4 show that the addition of fibers to water tends to decrease the gas holdup in a suspension at a given air flow rate (superficial gas velocity), presumably because of enhanced bubble coalescence (visual examination also indicated that larger bubbles formed).

In 2% pulp at high gas flow rates, Figure 4 shows that the gas holdup can exceed that of water, contrary to the results for other conditions. This may be an artifact due to enhanced channeling along the center of the column (a trend toward annular gas flow). The chord averaging of gamma densitometry in a circular geometry gives excess weight to void volumes at the center of the column. For example, a circular region at the center of the column with diameter D/2 will contribute to 50% of a gamma scan along that diameter, but will only contribute to 25% of the area average void volume. If the 2% suspension at high air velocity has most of the air flow occurring in the center of the column, it is possible that the diameter-averaged holdup could appear higher than the corresponding case in water if the water system had a more uniform distribution of gas. We have a number of column-averaged void volume measurements in the 10-cm column based on differential pressure transducer measurements for average column density, and while the results are not yet complete, the data to date indicate that the 2% pulp has a lower column-averaged holdup than water, making the results for 2% pulp in Figure 4 appear to be an indication of enhanced channeling or incipient annular flow.

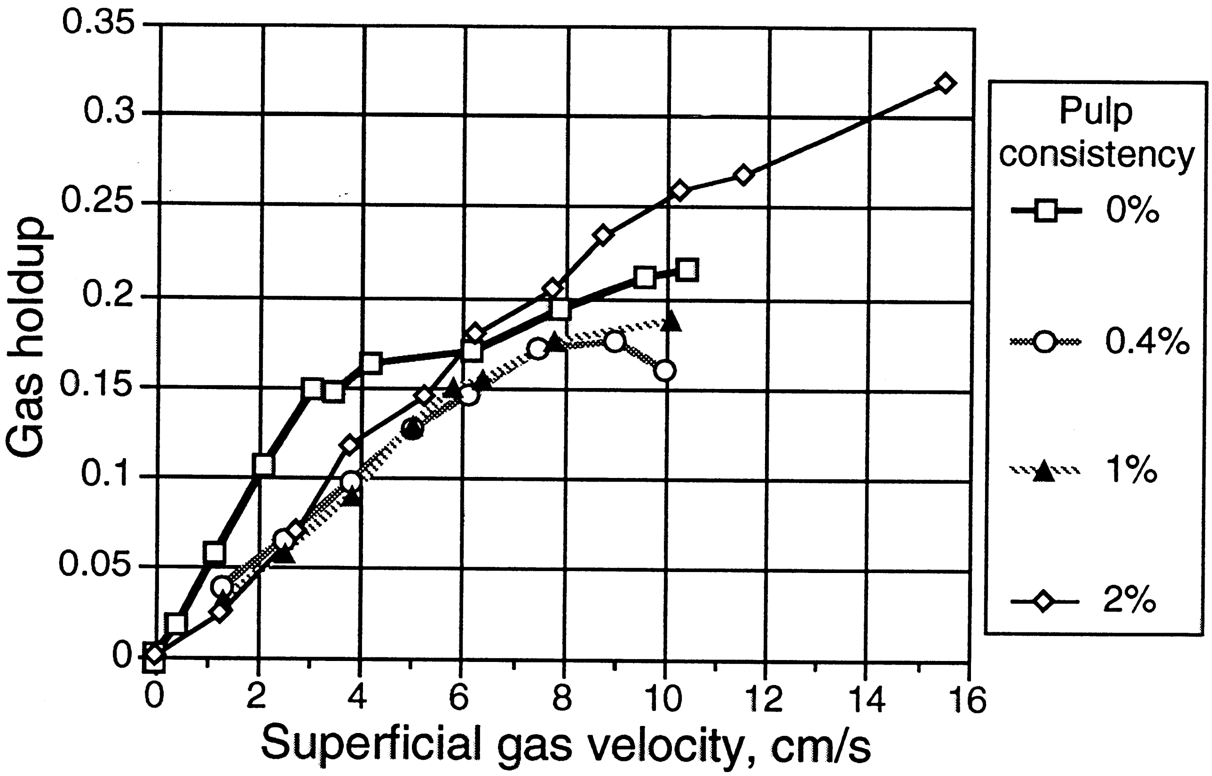

Figure 4. Diameter-averaged gas holdup at 22-24 cm above gas injection plane in the 10-cm stagnant pulp column for several consistencies, including pure water.

A variety of flow regimes were observed in these measurements. For low consistency suspensions, including water, at low gas flow rates, a bubbly flow regime exists in which individual bubbles (possibly with some early coalescence when fibers are present) rise through the suspension without significant interaction. The interaction of bubbles in the opaque pulp suspensions is difficult to determine, but a bubble flow regime at low velocities is likely (after some initial coalescence to allow the bubbles to rise through the flocs). A bubbly flow regime will generally show a linear relationship between air flow rate and holdup, with an intercept at the origin. As the gas velocity is further increased, the fluid motion takes on a churning nature in which large eddies of fluid (length scale approaching a column diameter) swirl in the column, often with distinct boundaries between one eddy and the next (fibers enhance the visibility of eddy boundaries). This may occur above 2-4 cm/s in the pulp suspension flows we examined. Bubbles tend to be larger, with some being on the order of 1 cm or larger. Fluid rising with gas near the center of the column (on the average) will tend to descend back down the column at the walls, though the motion is irregular due to the churning effect. At still higher gas velocities, the gas may form large bubbles or slugs (ca. 4 cm or more in diameter) or may form rapidly shifting columns or pathways with reduced surface area per unit volume. The surface becomes violent and significant amounts of pulp may be ejected.

Figures 5 through 9 show gas holdup at different column positions for individual pulp consistencies. The vertical positions shown represent only half of the positions scanned, with positions deleted only for clarity. In general, there is little change in holdup with vertical distance. The exceptions occur low in the column, at the first vertical position (6 cm above the gas injection plane) and near the top of the column. At the top of the column, the sloshing and splashing of the fluid can entrain additional air (especially small bubbles) that can increase the apparent holdup. Some of these entrained bubbles are taken downward in the recirculatory flow that descends along the column walls. The recirculation zone terminates near the first measurement position above the injection plane, resulting in less entrained air being returned to this position. Initially, we assumed that the low void volume generally seen near the bottom of the column was due to the air injectors being mounted at the wall. We assumed that the air was rising preferentially near the wall, not at the centerline, thus making gamma scans through the centerline appear to have unusually low void volume. However, measurement in the cocurrent flow system (discussed below) showed the same trends, with lower void volumes near the lower portion of the column. Recirculatory flow appears to provide the best explanation for the differing gas distribution in the vertical direction.

Figures 6 and 7 compare 0.4% pulp suspensions with and without surfactant. The surfactant was Triton X-100, added at a dosage of 1.7% on a dry fiber basis (200 μl per 3000 ml of pulp suspension). The surfactant has relatively little effect on the void volume distribution, possibly because coalescence mechanisms induced by flocs are more important than surface tension forces for these experiments. On the other hand, the surfactant may not have had a strong effect on surface tension since the Triton-free pulp had a surface tension of only 56 dyne/cm.

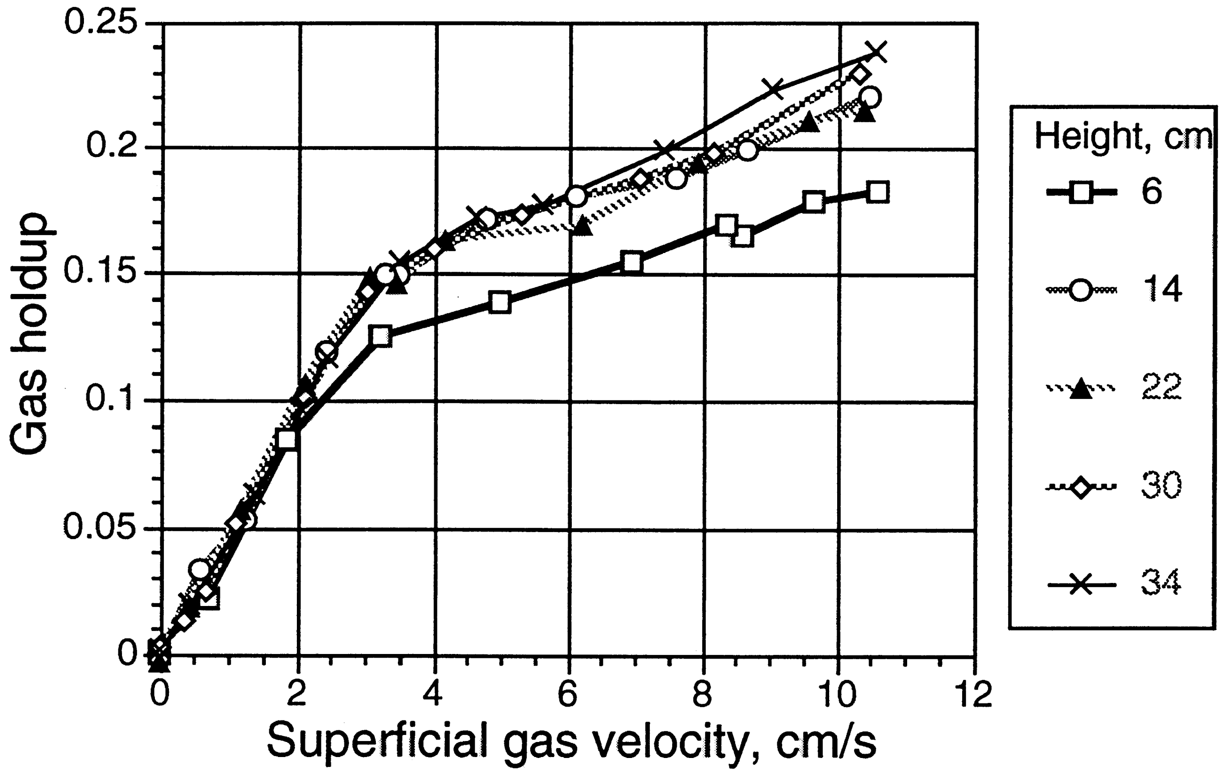

Figure 5. Centerline-averaged gas holdup in pure water, 10-cm column.

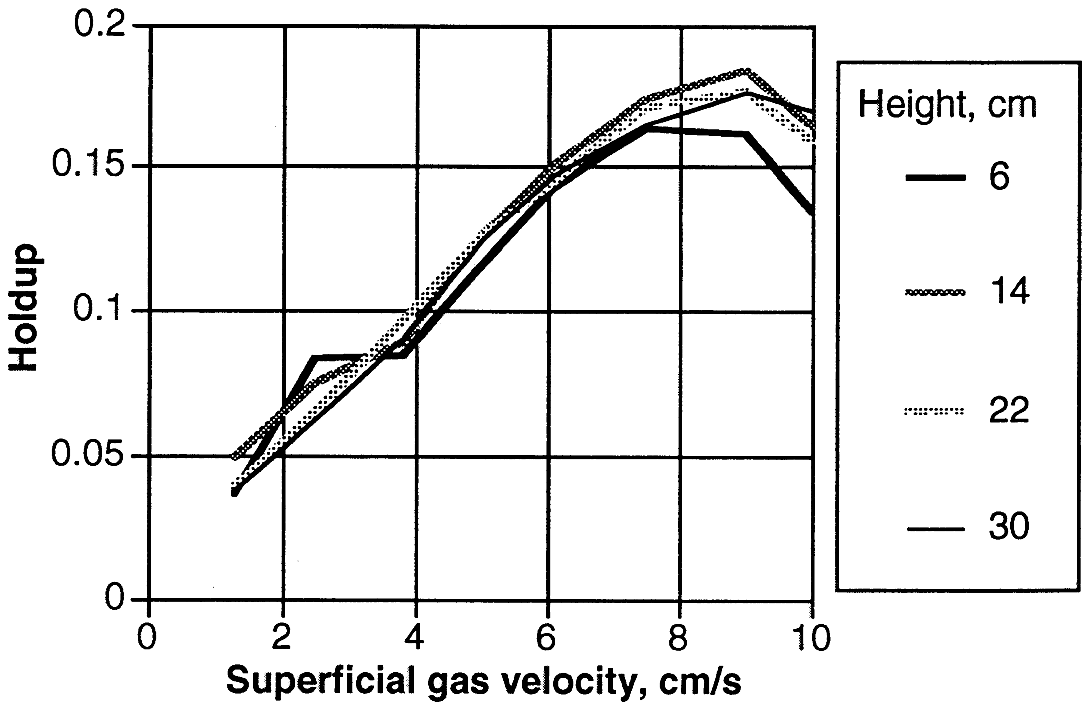

Figure 6. Diameter-averaged gas holdup in the 10-cm stagnant pulp column for several column positions (height above gas injection zone) with 0.4% pulp.

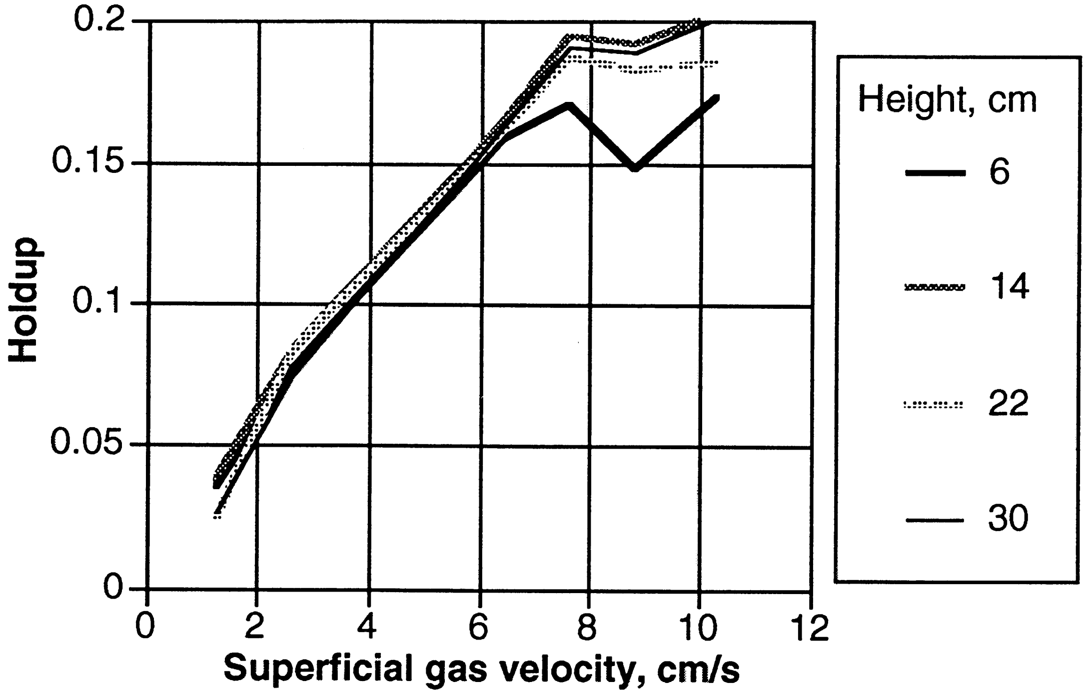

Figure 7. Diameter-averaged gas holdup in the 10-cm stagnant pulp column for several column positions (height above gas injection zone) with 0.4% pulp and added surfactant (1.7% Triton X-100 on a dry fiber basis).

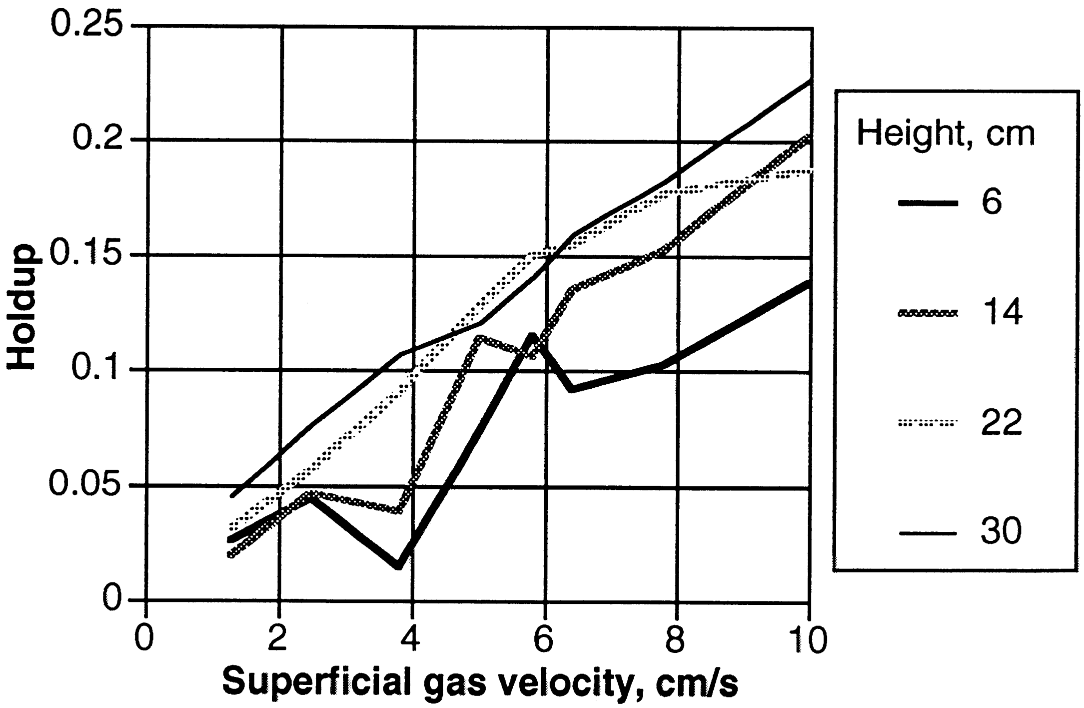

Figure 8. Diameter-averaged gas holdup in the 10-cm stagnant pulp column for several column positions (height above gas injection zone) with 1% pulp.

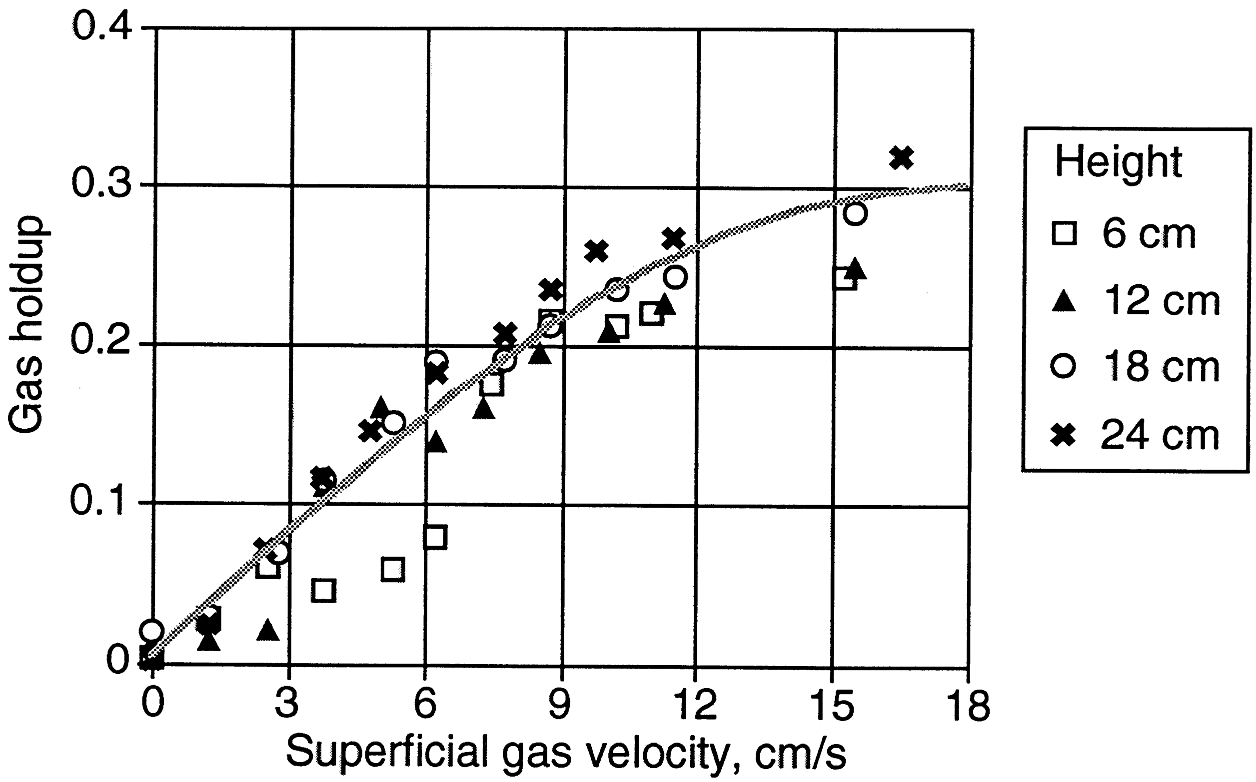

Figure 9. Diameter-averaged gas holdup in the 10-cm stagnant pulp column for several column positions (height above gas injection zone) with 2% pulp.

In Figures 6 and 7, the apparent decrease in gas holdup at high gas velocity (> 8 cm/s) is noteworthy. This may indicate a restructuring of the flow with less interfacial drag, possibly a transition toward enhanced channel flow as opposed to slug or churn flow. Figure 8 shows more variability in the measurements than typical, possibly due to problems in experimental methods. In Figure 9, a single curve is drawn through the available data. An important distinction between holdup curves in pulp (at least for consistencies above 0.4%) compared to water is that there is not a sharp break in the slope of the holdup curve as air flow is increased (e.g., compare Figure 9 with Figure 5). It may be that bubble coalescence induced by fibers forces the pulp holdup curves to begin with a lower slope, similar to the slope on the water curve after the transition away from a bubble flow regime. In other words, a true bubbly regime may be bypassed in pulp, with bubble-bubble interaction occurring long before a churn flow regime would have been reached in water.

Figures 10 through 13 show vertical profiles of gas holdup along the center of the 10-cm column for consistencies of 0.4% through 2%. In general, the variation of holdup with height is not dramatic, although the 1% suspension had significant gradients. When there is significant variation, the holdup at the bottom of the column is low. The holdup again increases as the turbulent, frothy top of the suspension is reached. Pulp consistencies of 1% or less (including water) showed the low holdup region for 5-10 cm above the gas injectors. Initially we assumed that this behavior was due to injection of gas at the wall of the cylinder, and that the gas at the wall had not had time to move toward the center of the column for measurements within 10 cm of the injection plane. However, similar results are obtained with the through-flow system, with average gas holdup lower near the bottom of the cylinder for the first 40 cm of the column above the conical mixing zone. In the cocurrent flow system, the jump in average holdup with height appears to be due to recirculatory flow bringing a portion of the air back down. The recirculation zone does not extend all the way down to the injection plane, so there is a rapid increase in gas holdup above the onset of large-scale recirculation.

Pronounced channel flow (still with churning and turbulence presence) occurs at very low air flow rates as the pulp consistency nears 2%. In addition, the presence of fibers can decrease the gas content in the pulp at moderate flow rates. Even though the fiber network system presents a barrier that slows the rise of an individual bubble, by the same token it holds bubbles in place until enough have coalesced to break through the network. Coalescence can lead to large bubbles that rise rapidly through the stock, rapidly enough to decrease the overall gas content compared to pure water or more dilute stock.

We also performed fiber length analysis on fibers retained by the froth and fibers retained in the pulp suspension. The Kajaani FS-100 results indicated that the froth preferentially removed long fibers.

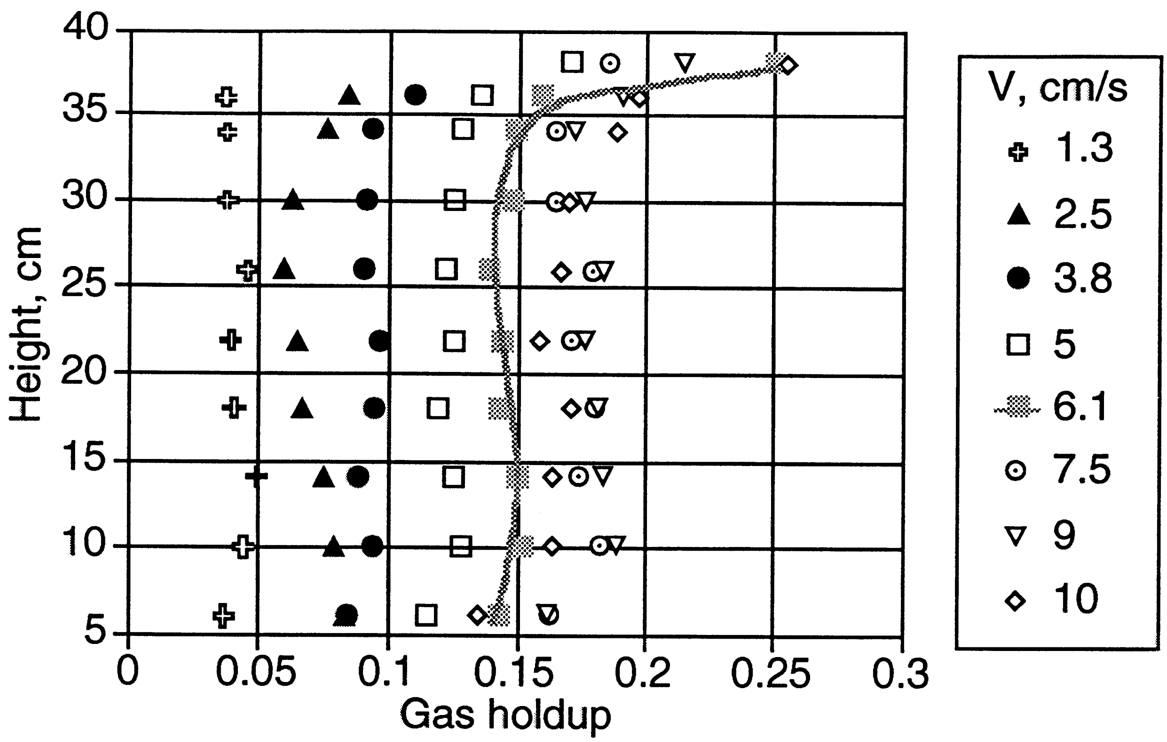

Figure 10. Vertical profiles of gas holdup in 0.4% pulp as a function of gas flow.

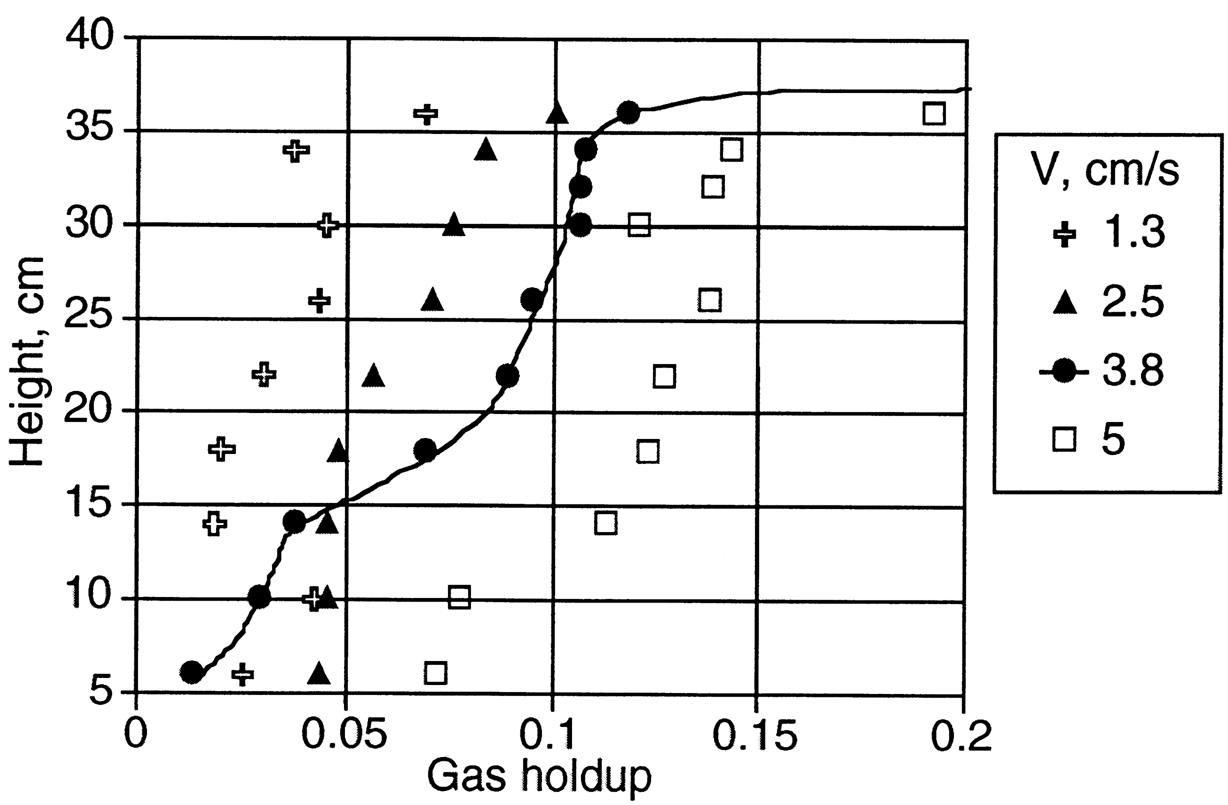

Figure 11. Vertical profiles of gas holdup in 1% pulp as a function of superficial gas velocity for low air flow rates.

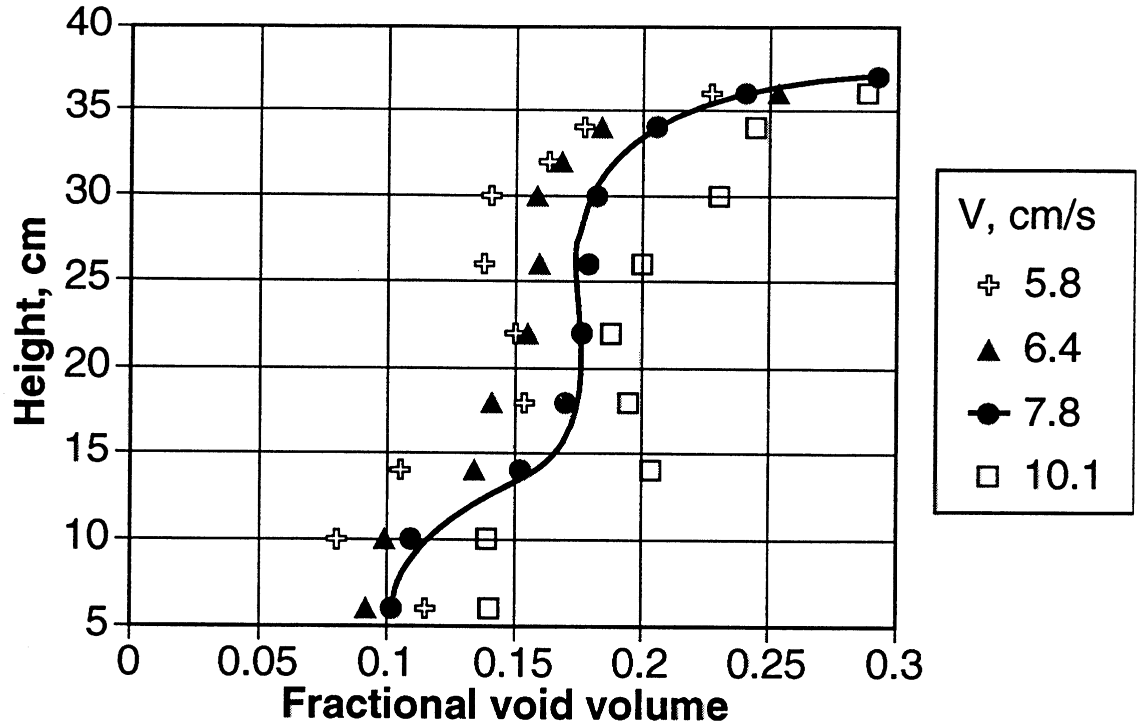

Figure 12. Vertical profiles of gas holdup in 1% pulp as a function of superficial gas velocity for high air flow rates.

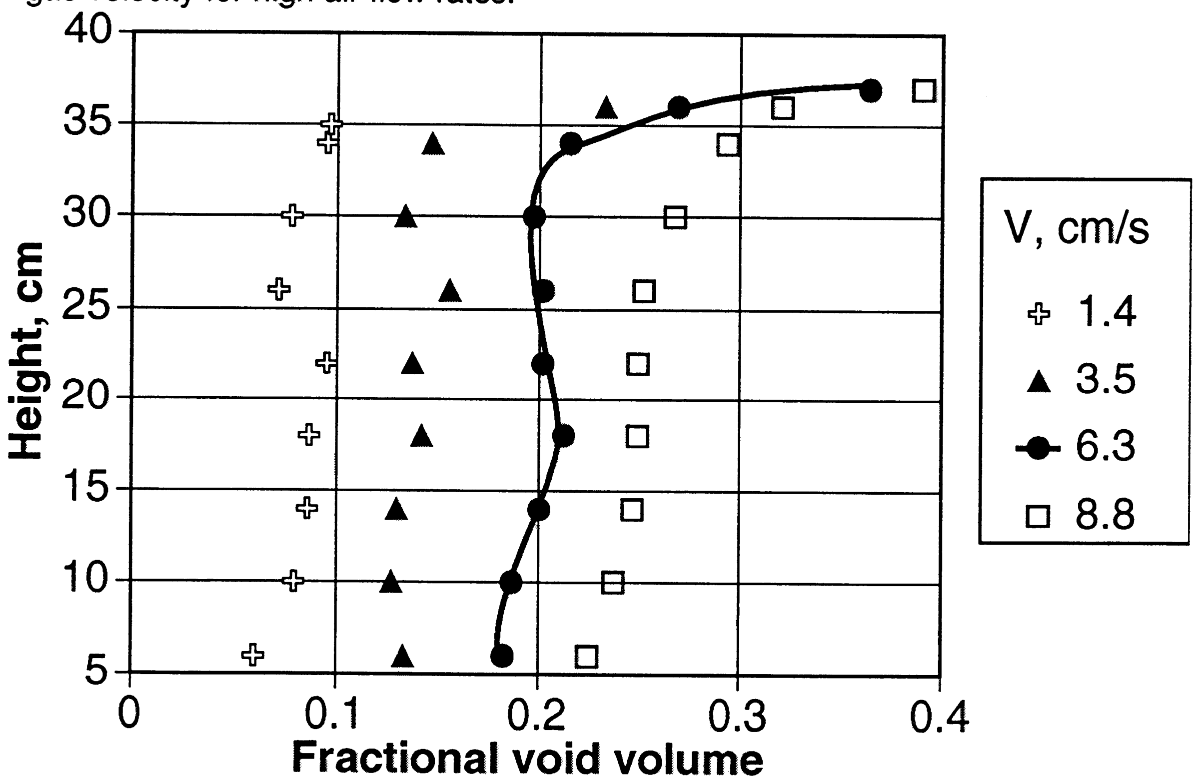

Figure 13. Vertical profiles of gas holdup in 2% pulp with added surfactant as a function of superficial gas velocity for high air flow rates.

Cocurrent Flow System

Unlike flow in the static column, the void volume of a 1% consistency pulp in the cocurrent flow system was generally found to be higher than the void volume in pure water at various combinations of liquid and gas flow rates. The reasons for this behavior are discussed later in this section. As a rule, flow regimes in the cocurrent flow column resembled those in the quiescent column, with bubbly flow in water at low gas velocities (roughly < 3 cm/s gas superficial velocity) followed by flow with a churn-turbulent nature at higher gas velocities. In pulp, the flow had a churn-turbulent nature at lower velocities, though only the outer portion of the flow could be seen (true for the quiescent column as well). What appeared to be churn-turbulent flow in some cases may have simply been in a transition regime from ideal bubble to churn-turbulent flow, for the gas holdup curves (holdup versus gas velocity) for pulp and for water were linear over the limited range of gas flow rates used. However, in pulp flows at the higher gas flow rates, some large bubbles were visible, with diameters on the order of 2-5 cm, which created large, swirling eddies in the flow as they rose, indicative of churn-turbulent flow. An improved tank system is being installed to allow higher air injection rates in the future. Precise identification of flow regimes in the future will require examination of transient differential pressure measurements across the column to obtain information not available by visual inspection.

Figures 13 through 17 show the average void volume at different heights for different superficial velocity combinations in a 1% ONP suspension and in water. The results are similar to those found for the stagnant column. At low air velocities, there is very little change in average void volume at the different heights. However, at higher air superficial velocities, the average void volume increases with the height of the column. The increased void volume can be attributed to the gas being forced to the center of the column by the air rich pulp that is recirculating due to the constricting end of the through-flow unit.

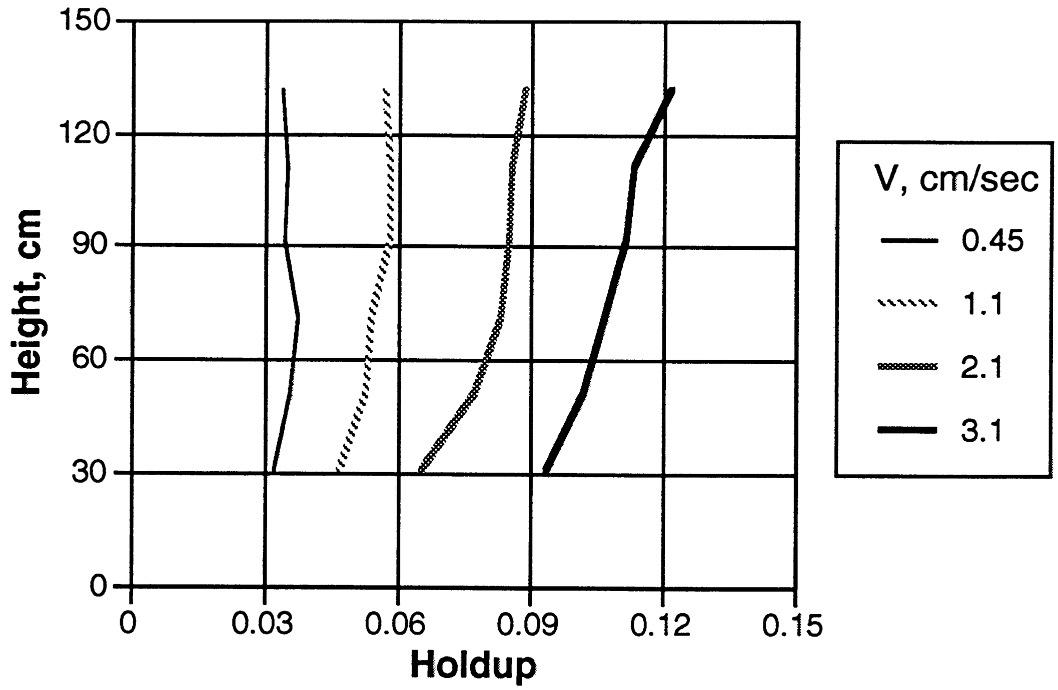

Figure 13. Cross-section averaged gas holdup in the cocurrent flow column with a liquid flux of 2.5 cm/s (5 gpm flow rate) and a 1% ONP suspension.

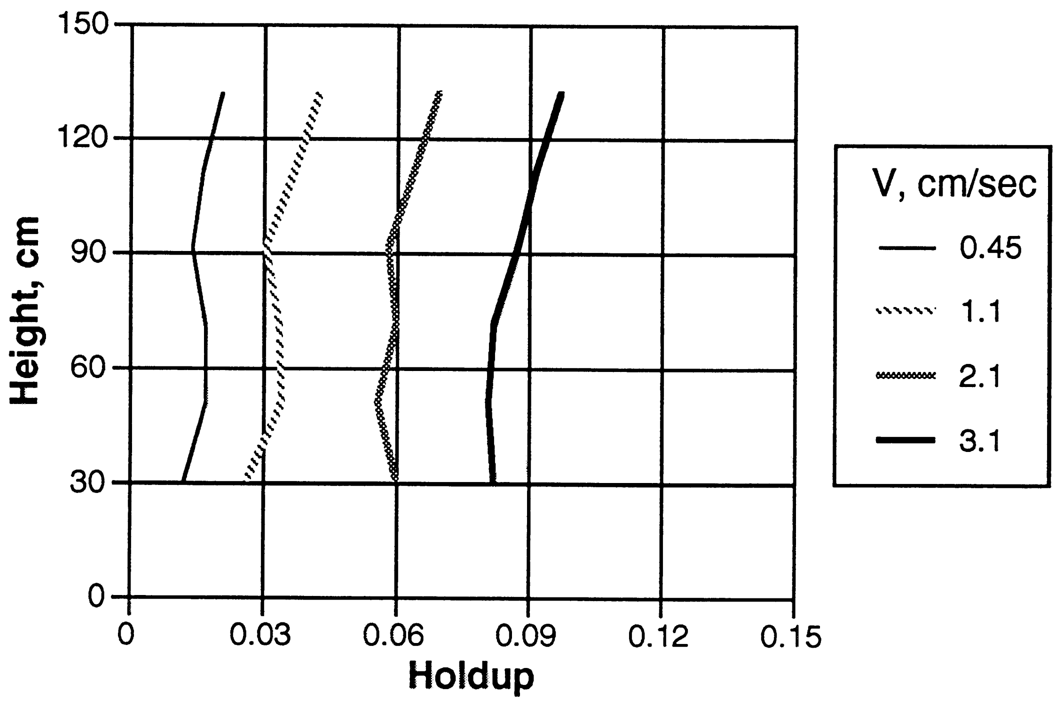

Figure 14. Cross-section averaged gas holdup in the cocurrent flow column with a liquid flux of 2.5 cm/s (5 gpm flow rate) in clear water.

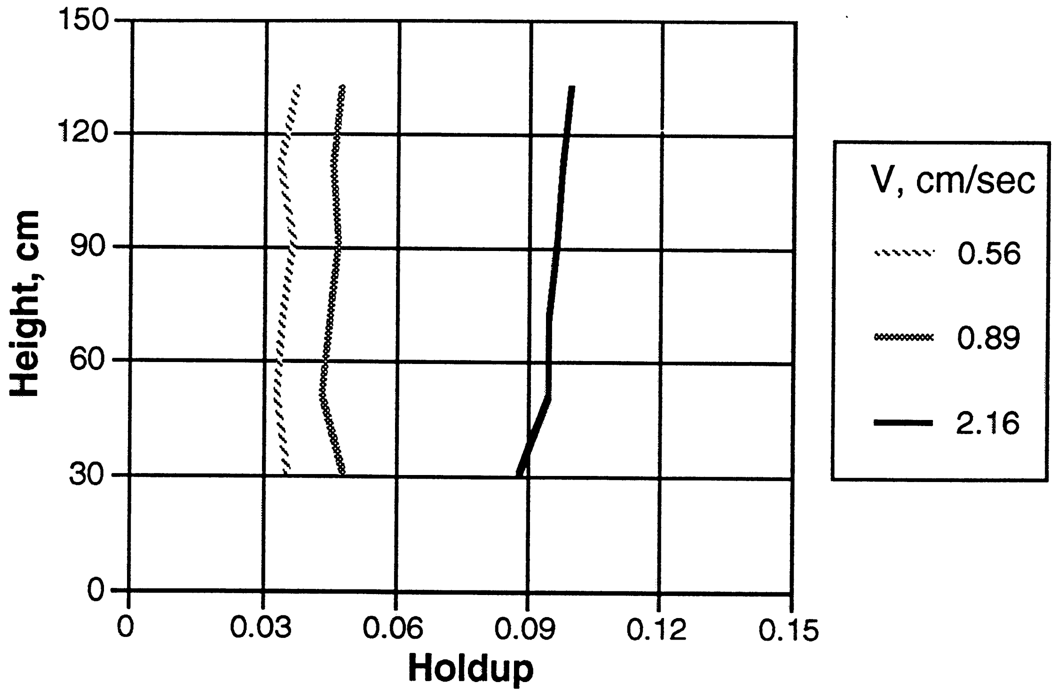

Figure 15. Cross-section averaged gas holdup in the cocurrent flow column with a liquid flux of 5.05 cm/s (10 gpm flow rate) and a 1% ONP suspension.

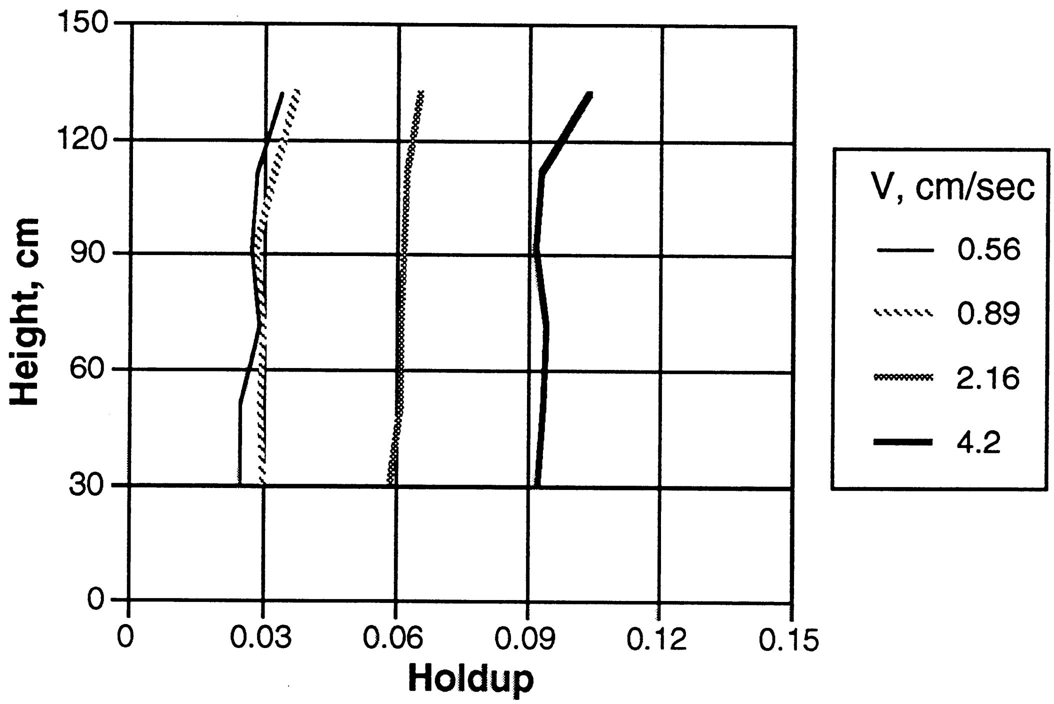

Figure 16. Cross-section averaged gas holdup in the cocurrent flow column with a liquid flux of 5.1 cm/s (10 gpm flow rate) in clear water.

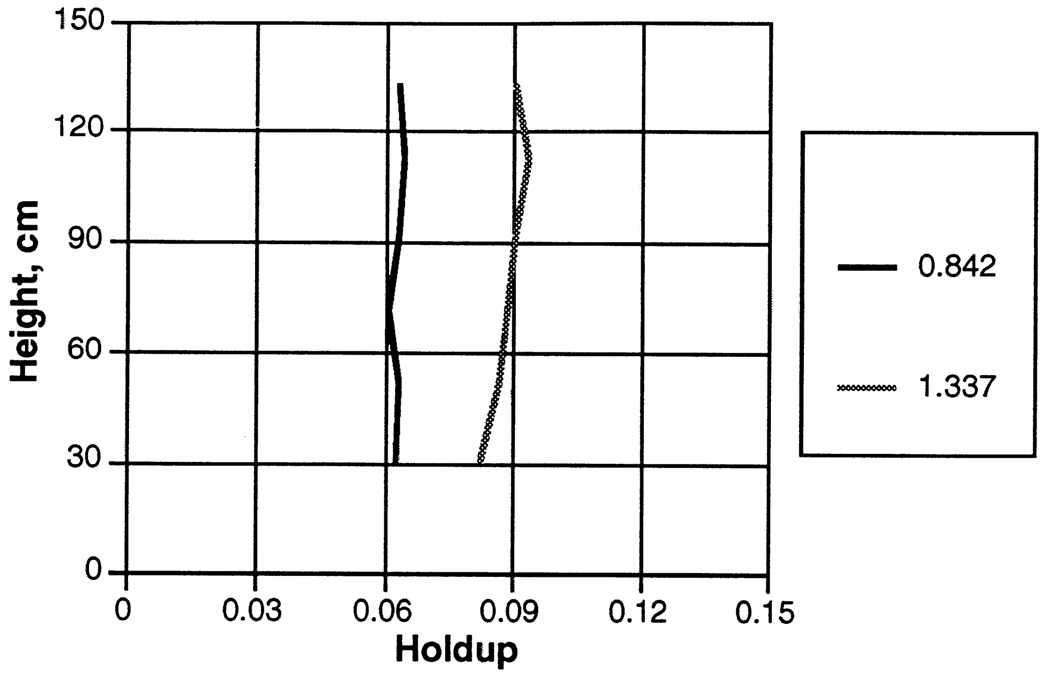

Figure 17. Cross-section averaged gas holdup in the cocurrent flow column with a liquid flux of 7.5 cm/s (15 gpm flow rate) and a 1% ONP suspension.

Comparison of pulp and water results for gas holdup in the cocurrent flow system shows that the gas holdup has increased, in most cases, by the addition of fibers. This is shown more clearly by considering the column averages (holdup values averaged over cross sectional area and over column height) given in Figures 18-20. Part of the increased gas content in the pulp flow may be due to entrainment of air in the pulp as it flows from the holding tank back to the pump. For example, extrapolation in Figure 18 of the pulp curve to an injected gas flux of zero gives a holdup of 2%, indicating that 2% air content was present in the pulp as it left the holding tank. If so, the difference between the pulp and water results in Figure 18 may be due primarily to extra air that was present in the pulp before passing through the air injector. However, at higher liquid flow rates (Figures 19 and 20), the higher holdup in the pulp relative to water cannot be explained by higher initial air loading alone (in fact, the extrapolated air content at zero injected gas flux is about 2% for both water and pulp suspensions at the two higher pulp flow rates). In Figures 19 and 20, the liquid superficial velocity is greater than the gas superficial velocity for all measurements (higher air flow rates increased pressure drop and increased air accumulation in the pump, hindering pump operation). Under these conditions, the dispersed bubbles, slowed in their ascent by the fibers, can be carried away by the pulp flow before other bubbles rise and coalesce with them. It seems that fibers cause coalescence when small bubbles cannot break through the network rapidly enough to avoid being hit by subsequent bubbles, and that aggregation continues until a large bubble has enough buoyant force to escape. When the pulp itself is flowing upwards as well, bubbles can avoid collisions by being carried away before later bubbles arrive.

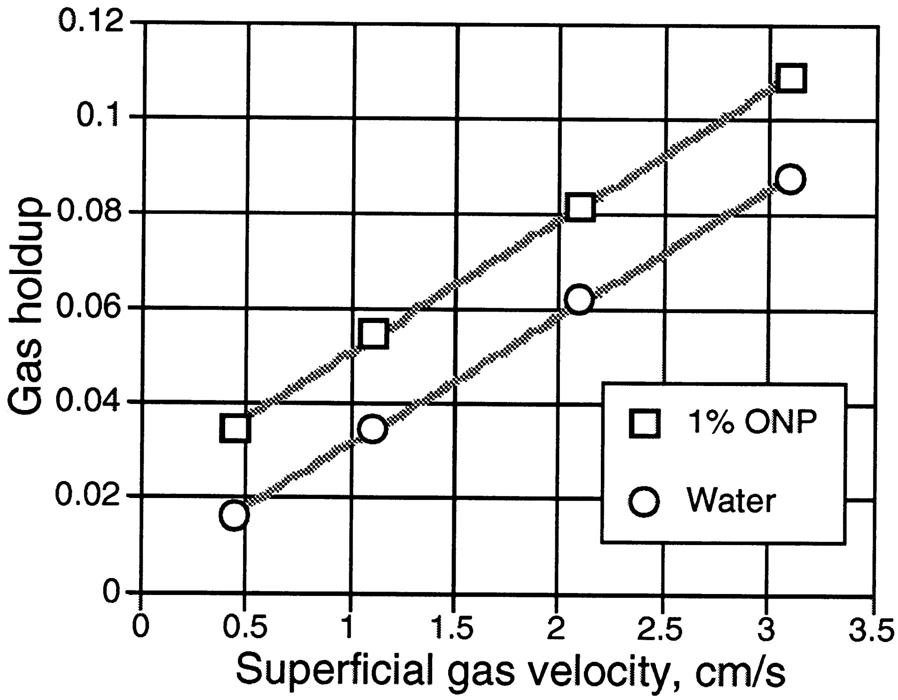

Figure 18. Column averaged gas holdup in 1% ONP and in water as a function of gas flux for a constant liquid flux of 2.5 cm/s (5 gpm).

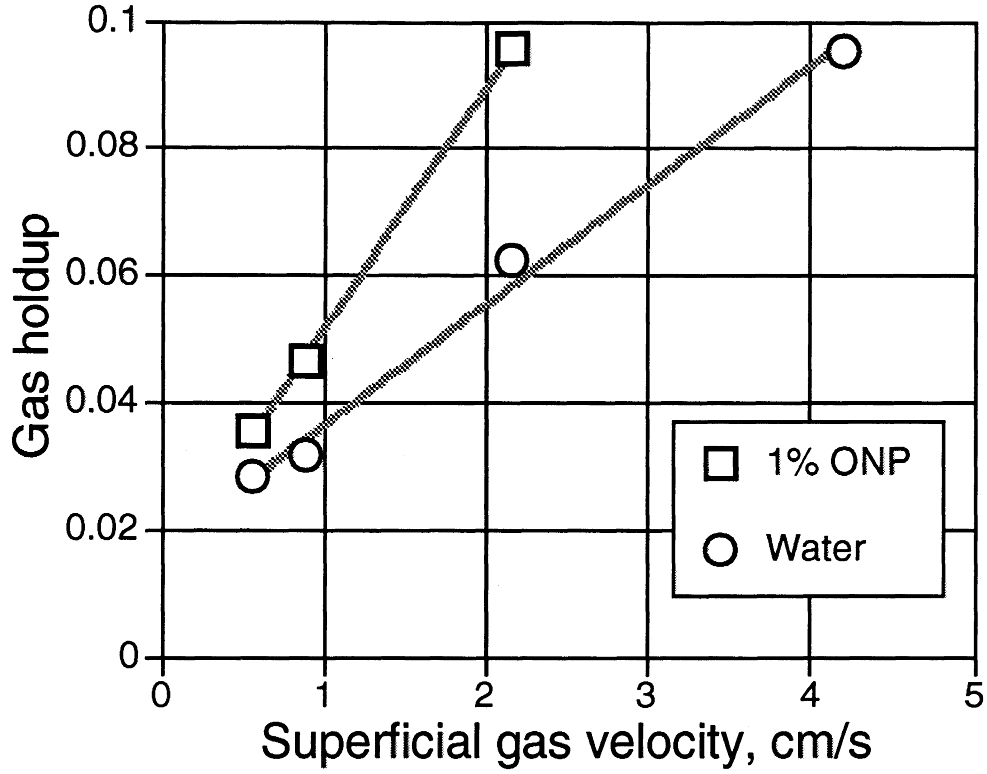

Figure 19. Column averaged gas holdup in 1% ONP and in water as a function of gas flux for a constant liquid flux of 5.1 cm/s (10 gpm).

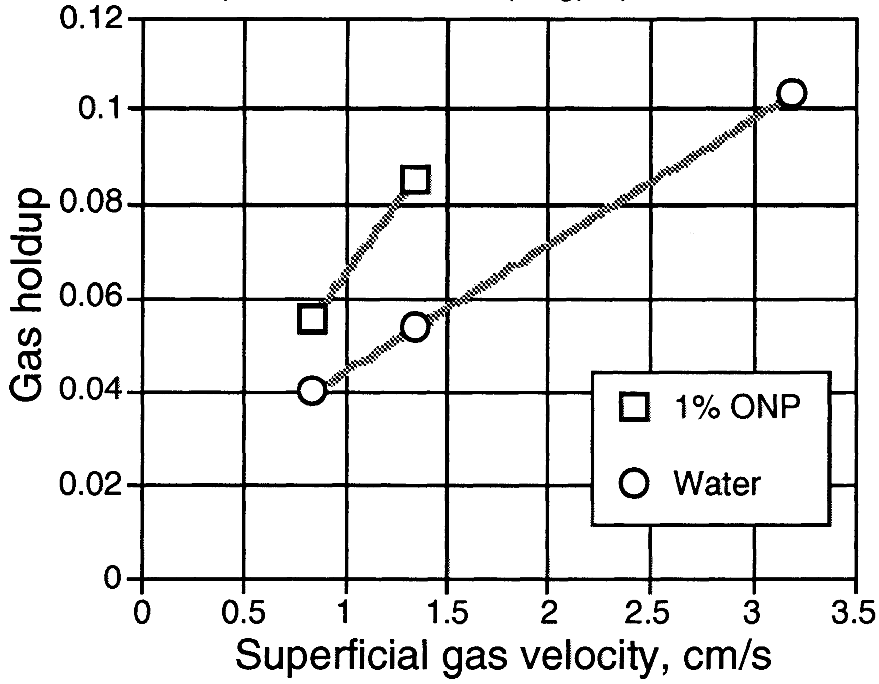

Figure 20. Column averaged gas holdup in 1% ONP and in water as a function of gas flux for a constant liquid flux of 7.6 cm/s (15 gpm).

Figures 21 through 35 show the radial and vertical variations observed in the chord averaged gamma densitometry measurements through the cocurrent column for both pulp (1% ONP) and water at several liquid and gas fluxes. Both water and pulp results are similar, probably due to the decreased fiber-induced coalescence in the cocurrent flow. Flows at lower liquid velocity are more likely to have an uneven gas distribution in the lower portion of the column, probably due to the Coanda effect, causing the incoming flow to preferentially follow the wall of the conical expansion in some cases, or to oscillate from one side to the other. The gas distribution tends to be reasonably symmetric about the centerline after 80 cm of upward travel. The radial distribution of the chord-averaged void volume appears parabolic, commonly having a centerline average holdup two or three times the values at the wall. Since the centerline chord-average includes portions near the wall as well as at the center of the column, one might infer that the actual holdup at the center of the column is more than two or three times as great as the holdup at the walls. Using the assumption of radial symmetry, the chord averages at various radial locations can be used to solve for the true local radial holdup distribution that would yield the measured chord averages. Unfortunately, the inverse method (a matrix inversion) is relatively unstable, meaning that small changes in the measured chord averages yield large changes in the symmetric radial distribution. Heavy smoothing of the actual data is usually required to avoid negative values in the radial solution. Where reasonable results were obtained, the calculated center gas holdup generally was three to seven times as great as the values near the wall, but this approach is questionable since most of our measurement show some departure from true radial symmetry.

Figure 21 provides an example for discussion. A superficial gas velocity of 3.1 cm/s flows in water having a superficial velocity of 2.5 cm/s. The air rising through the conical inlet probably fluctuates from side to side, yielding an apparent twin-peaked distribution in the time-averaged measurements near the inlet. The gas distribution becomes more symmetrical as the flow rises through the column. At 130 cm above the inlet, the radial distribution of chordal averages is nearly parabolic and on center. Chord-averaged centerline gas content is about twice that at the walls.

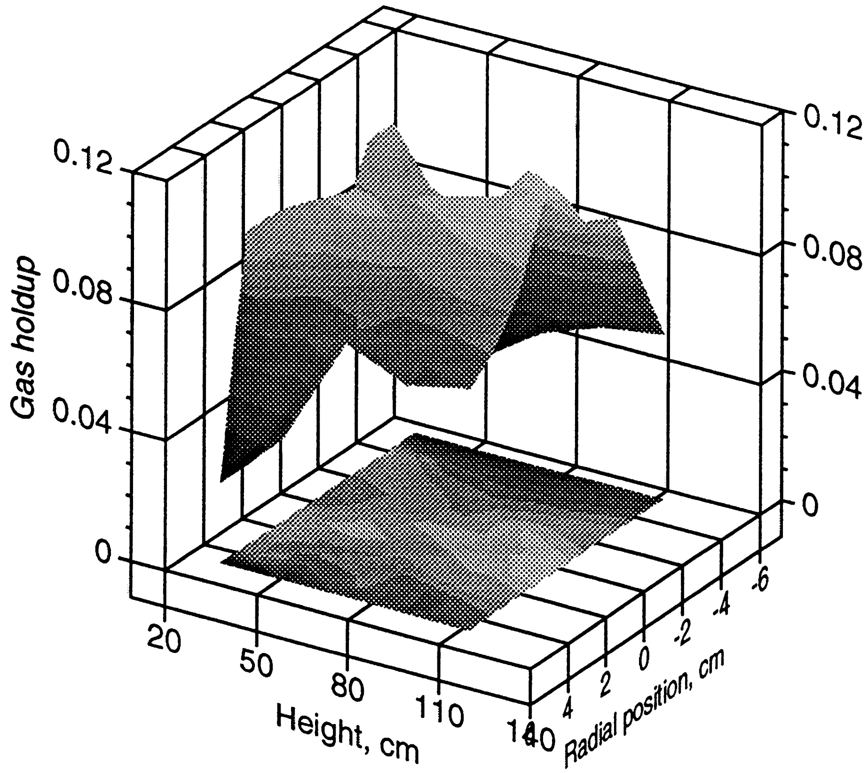

Figure 21. Chord-averaged void volume in the cocurrent flow system for water with superficial velocities of 2.5 cm/s for water and 3.1 cm/s for air.

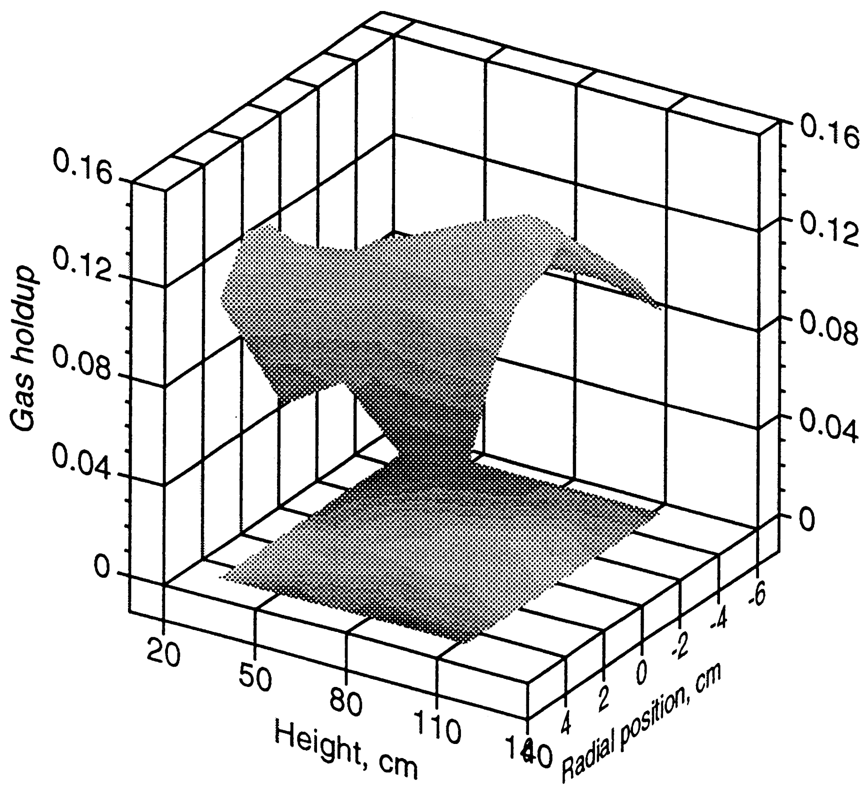

Figure 22. Chord-averaged void volume in the cocurrent flow system for 1% ONP with superficial velocities of 2.5 cm/s for pulp and 3.1 cm/s for air.

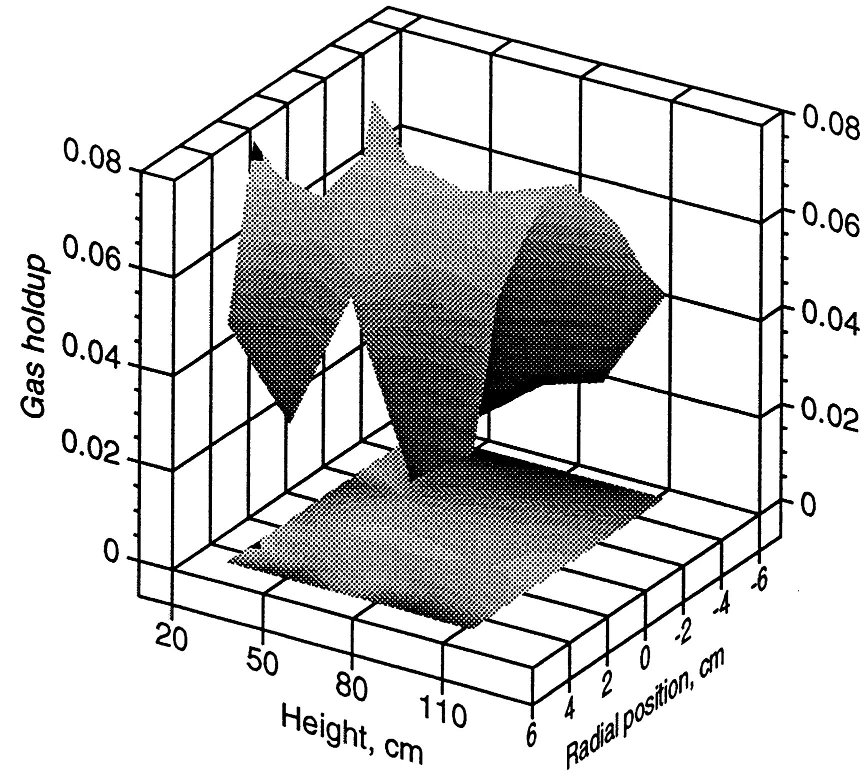

Figure 23. Chord-averaged void volume in the cocurrent flow system for water with superficial velocities of 2.5 cm/s for water and 2.1 cm/s for air.

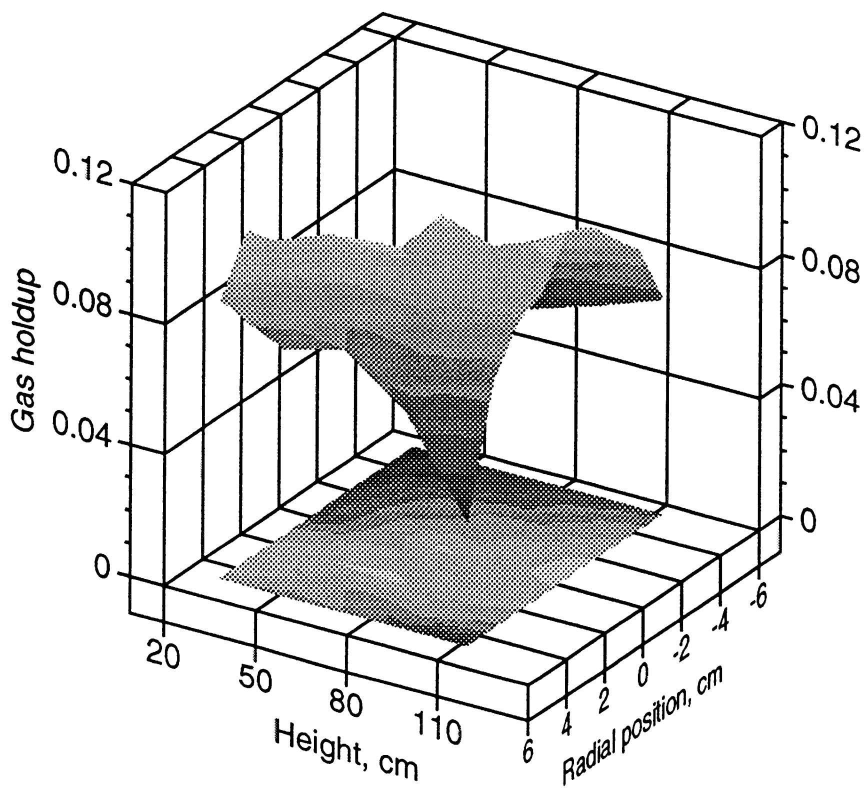

Figure 24. Chord-averaged void volume in the cocurrent flow system for 1% ONP with superficial velocities of 2.5 cm/s for pulp and 2.1 cm/s for air.

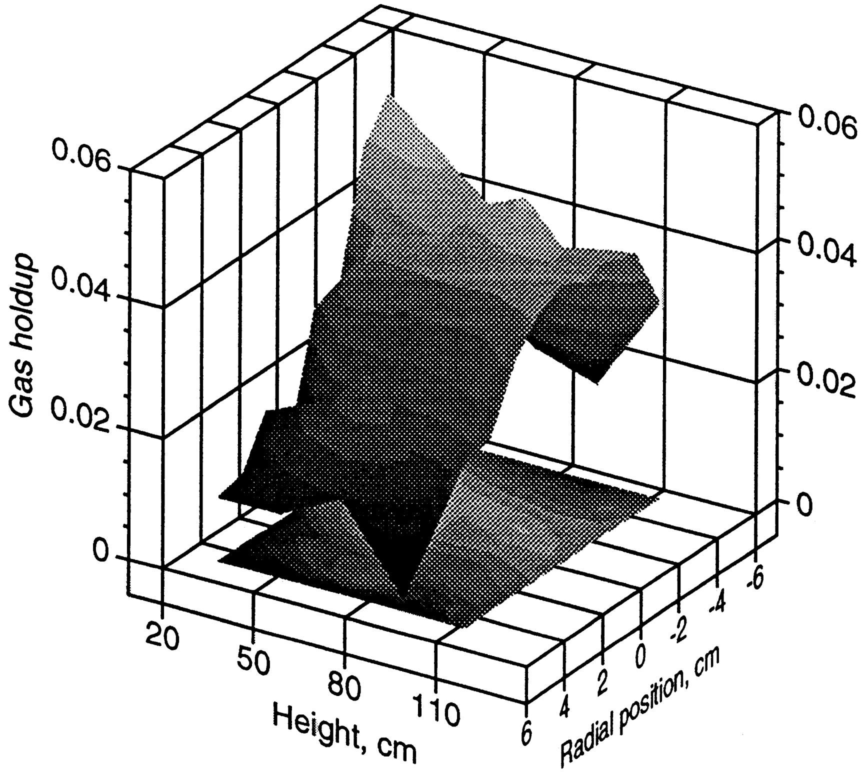

Figure 25. Chord-averaged void volume in the cocurrent flow system for water with superficial velocities of 2.5 cm/s for water and 1.1 cm/s for air.

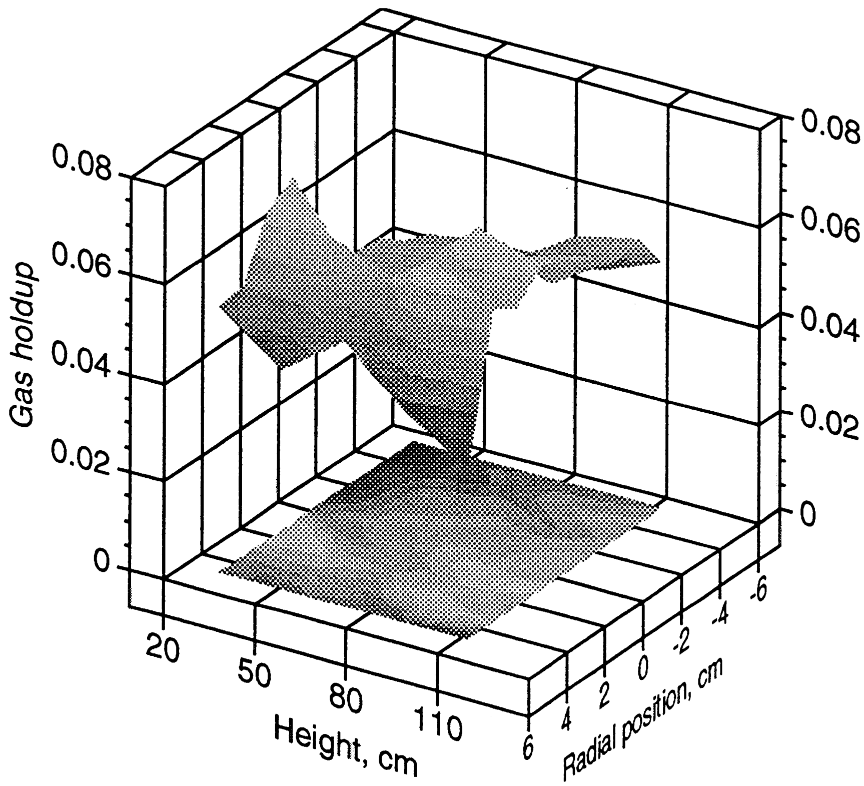

Figure 26. Chord-averaged void volume in the cocurrent flow system for 1% ONP with superficial velocities of 2.5 cm/s for pulp and 1.1 cm/s for air.

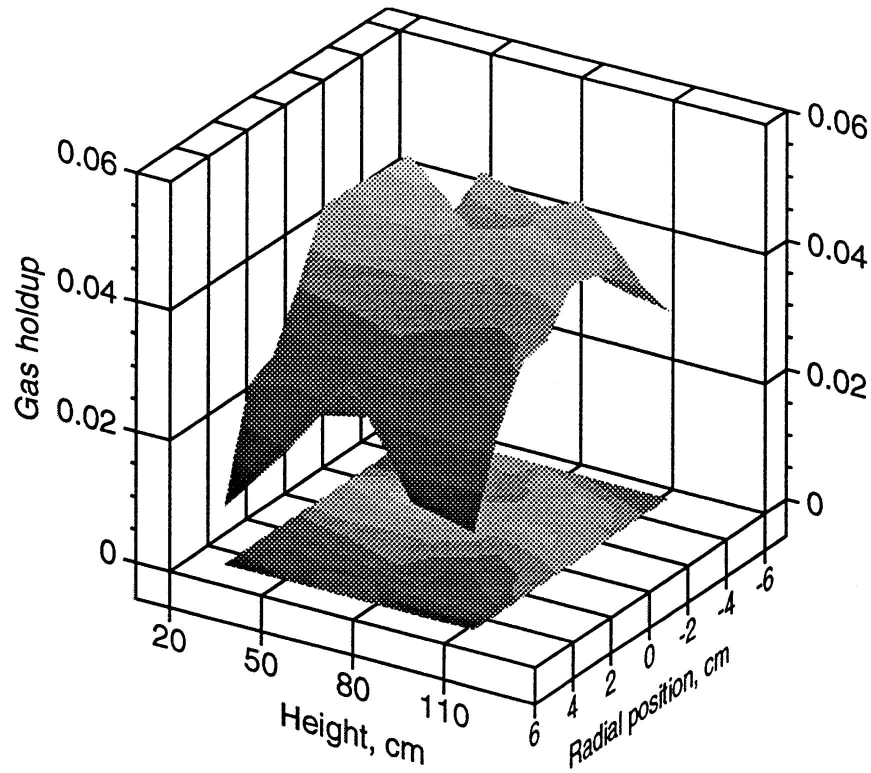

Figure 27. Chord-averaged void volume in the cocurrent flow system for 1% ONP with superficial velocities of 2.5 cm/s for pulp and 0.45 cm/s for air.

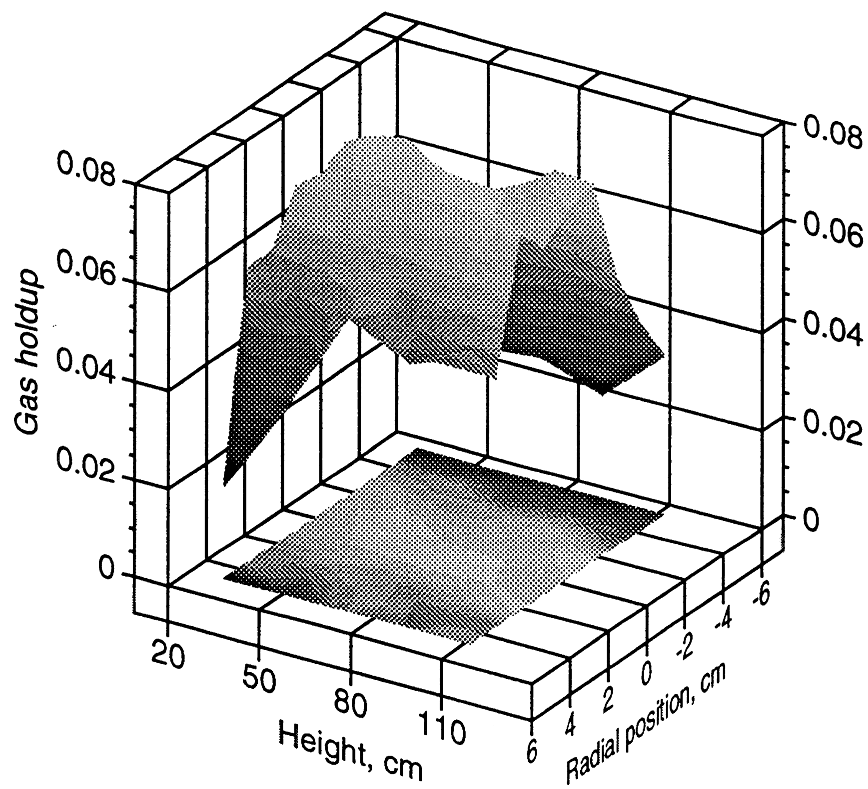

Figure 28. Chord-averaged void volume in the cocurrent flow system for water with superficial velocities of 5.1 cm/s for water and 0.89 cm/s for air.

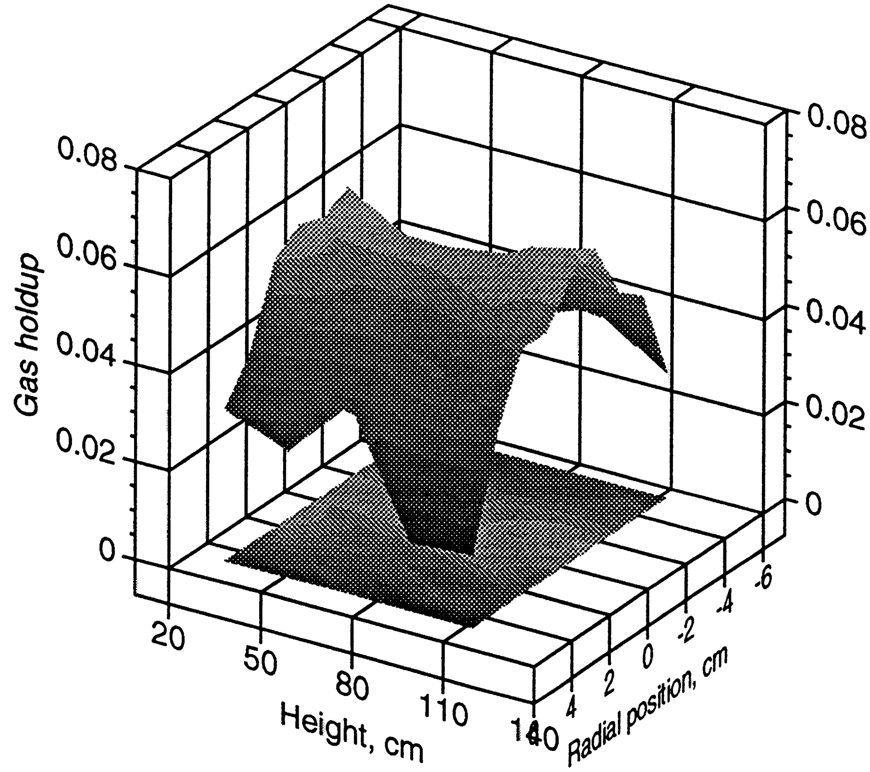

Figure 29. Chord-averaged void volume in the cocurrent flow system for 1% ONP with superficial velocities of 5.1 cm/s for pulp and 0.89 cm/s for air.

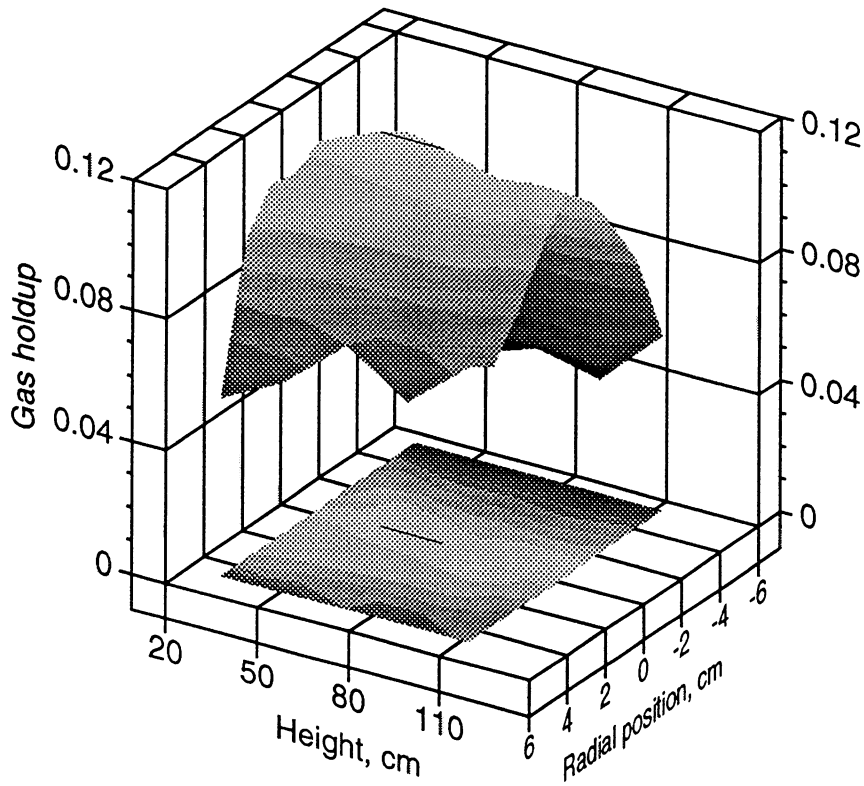

Figure 30. Chord-averaged void volume in the cocurrent flow system for water with superficial velocities of 5.1 cm/s for water and 2.16 cm/s for air.

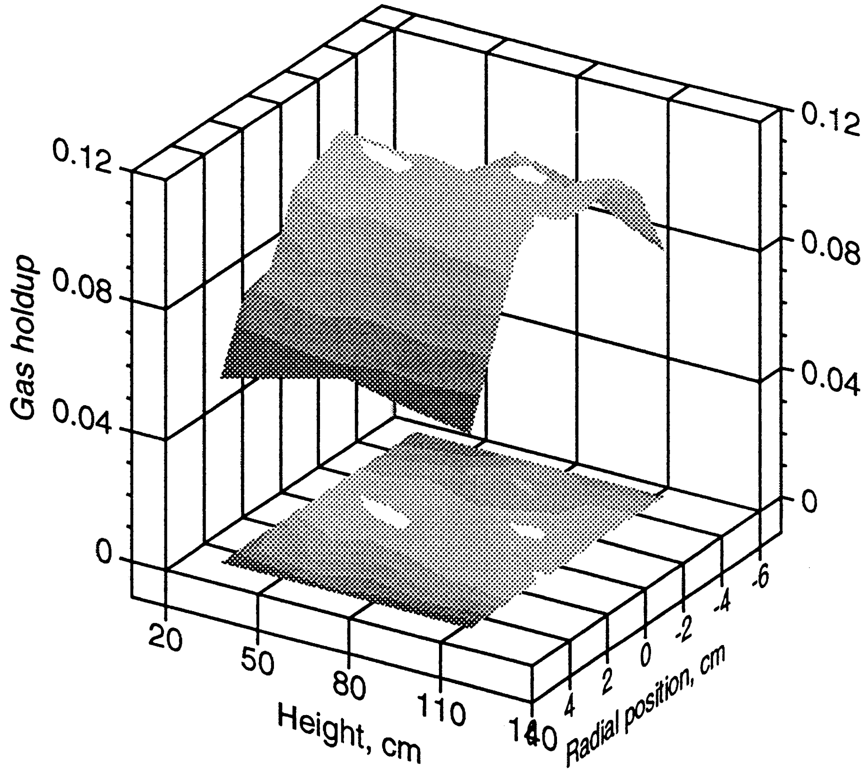

Figure 31. Chord-averaged void volume in the cocurrent flow system for 1% ONP with superficial velocities of 5.1 cm/s for pulp and 2.16 cm/s for air.

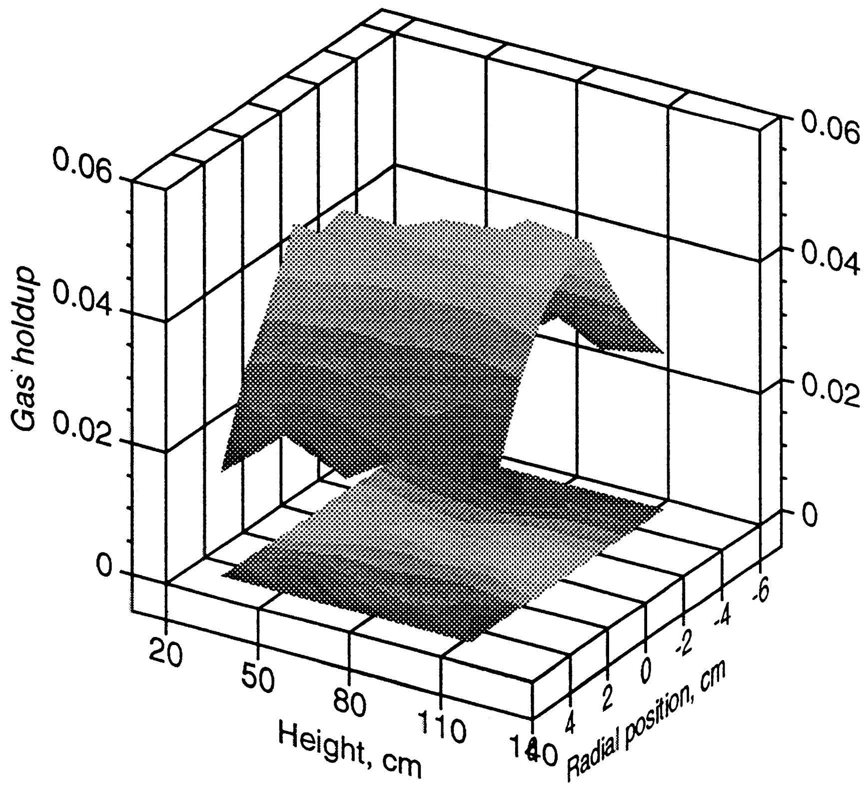

Figure 32. Chord-averaged void volume in the cocurrent flow system for 1% ONP with superficial velocities of 5.1 cm/s for pulp and 0.56 cm/s for air.

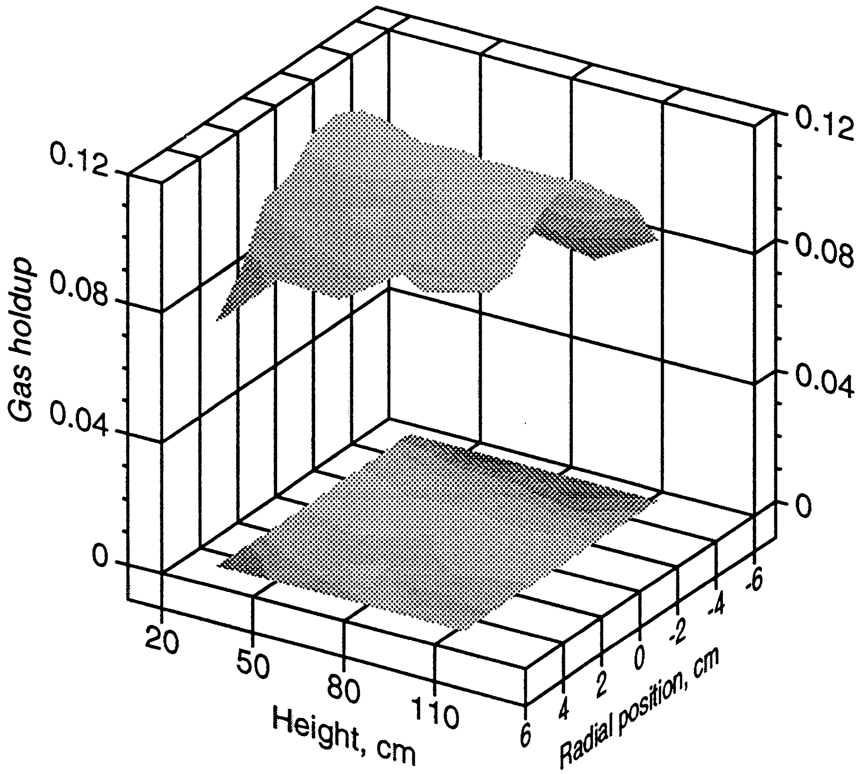

Figure 33. Chord-averaged void volume in the cocurrent flow system for water with superficial velocities of 7.6 cm/s for water and 1.34 cm/s for air.

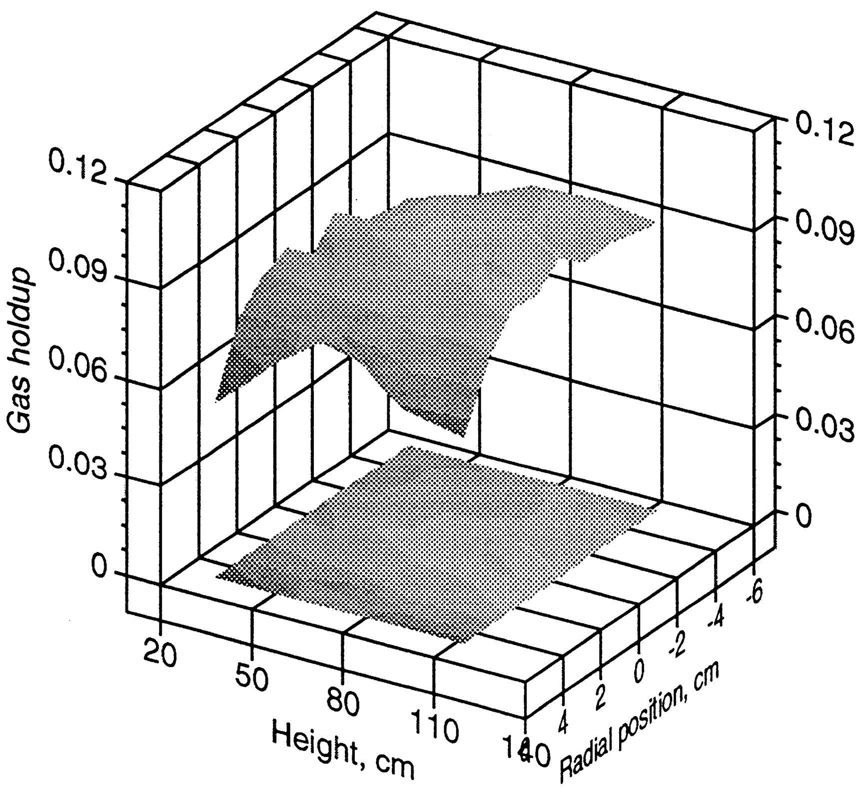

Figure 34. Chord-averaged void volume in the cocurrent flow system for 1% ONP with superficial velocities of 7.6 cm/s for pulp and 1.34 cm/s for air.

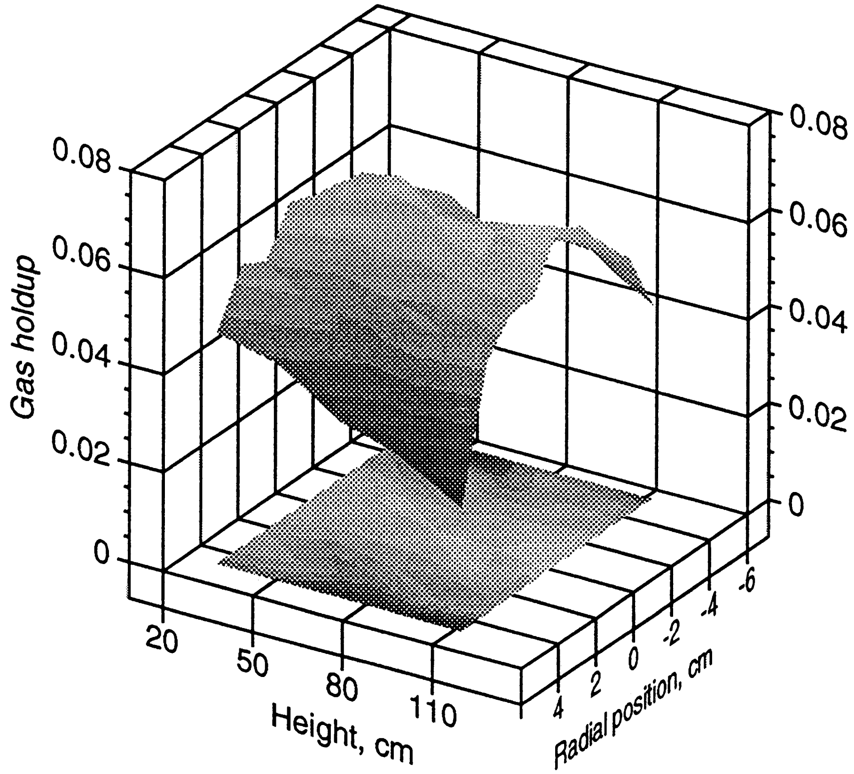

Figure 35. Chord-averaged void volume in the cocurrent flow system for 1% ONP with superficial velocities of 7.6 cm/s for pulp and 0.89 cm/s for air.

DISCUSSION AND CONCLUSIONS

These experiments have shown that the presence of fibers in a quiescent slurry promotes bubble coalescence. Increased bubble coalescence reduces the total surface area of the bubbles in the air-pulp mixture which can lower ink removal efficiencies in flotation units. In addition, increased bubble coalescence can lead nonuniform gas distribution (ultimately channel flow). Liquid recirculation is an important factor, driven by the buoyant flow of the gas that migrates away from the walls. Channel flow occurs above a critical gas velocity that depends on pulp consistency.

In the cocurrent flow system, fibers can increase the gas holdup when the pulp suspension is flux is sufficient to carry away bubbles before they coalesce with subsequent bubbles and break through the network that impedes their rise. The impediment to buoyant flow provided by the network thus increases the average gas content. Our current flow results have been limited by equipment restrictions, making our range of gas fluxes limited to 4.5 cm/s or less. Higher fluxes are needed to better simulate the types of flow that may occur in commercial deinking systems.

Cocurrent flow also exhibited flow recirculation, with gas and fluid rising more rapidly in the enter and flowing down at the walls. Recirculation may have been enhanced by the conical restriction at the top of the column (from a diameter of 12.7 cm down to 5 cm, the diameter of the return line to the holding tank). We will soon modify our flow system to have an open overflow region at the top. Recirculation will still be expected, however.

The presence of large bubbles in a commercial flotation system due to either strong churn-turbulent flow, slug flow, or local channel flow, may cause the surface of the system to become agitated in the regions where large bubbles or large streams of air are preferentially rising. This may be what some developers refer to as "detrimental turbulence" at high air flow rates (17), which is said to cause a drop in ink removal. We suspect the problem is not turbulence *per se*, but low interfacial area.

The mechanisms that lead to large bubbles and the general mechanism of coalescence due to fibers when the superficial gas velocity is greater than the liquid velocity suggests that the initial bubble size distribution entering a flow system does not persist. The actual bubble size distribution in a flotation deinking cell may be only slightly linked to the original bubble size distribution. There is a need to develop techniques for determining the bubble size distribution in pulp slurries, and student work is currently in progress on this important topic.

The recirculation observed in both flow systems reflects the importance of geometry in a flotation unit. We expect more confined geometries would inhibit recirculatory flows. Wall effects (viscous shear) impede recirculation, for high shear would be required to have upward flow at the center and downward flow at the walls. Instead, wall effects promote a more uniform flow and can be expected to increase the gas holdup and the interfacial area.

These early results indicate the need for more understanding of the flotation deinking process. Better understanding would aid in the design and operation of deinking cells that would create a more efficient deinking process.

PLANNED WORK

In order to promote improvements in flotation deinking processes, we need to fill the serious void in knowledge concerning gas-liquid-solid transport mechanisms in flotation deinking processes. A three-pronged approach will be pursued, combining macroscopic characterization of flow regimes (phase mapping), special effects studies focusing on specific mechanisms, and modeling to integrate experimental information and to identify key transport parameters and equations in three-phase fibrous systems. Specific tasks are:

A. Characterize the macroscopic hydrodynamics of multiphase fibrous suspensions relevant to flotation deinking. Determine the flow regimes that occur in gas-water-fiber systems relevant to flotation deinking. Create phase diagrams showing the regimes (e.g., churn flow, bubbly flow, channel flow) as functions of gas flow rates, fiber consistency, liquid flow, etc. Determine specific mechanisms that affect bubble dynamics in a fibrous suspension. For example, we will see how fibers prevent or promote coalescence in turbulent flow, and relate the behavior to fiber properties. The impact of fibers on bubble size and interfacial area is of great importance, and will be investigated using a variety of tools including flash x-ray radiography.

B. Examine the benefits of improved geometry and flow modification for enhanced ink removal. For example, our preliminary work shows that channel flow of gas in the pulp is likely when pulp consistency exceeds 1.5% or when gas flow rates become high. Increased consistency would be very desirable in reducing liquid effluent from deinking operations. Pulsed jets or other transitory flow features may be useful in preventing column formation and maintaining good interfacial area at elevated consistency without the need for mechanical agitation throughout the system (system-wide mechanical agitation leads to backmixing that can decrease the efficiency of the flotation cell in much the same way that a continuous stirred tank reactor is less efficient than a plug flow reactor for a given reactor volume for a positive-order reaction). We have also found evidence that smaller tank diameters may lead to better gas flow regimes (less channeling), suggesting that large cells could be improved with retrofitted vertical baffles or other structures to break the flow into smaller cells. These concepts will be tested in the near future.

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LITERATURE CITED

1. Wallis, G. B., "One-dimensional Two-phase Flow," McGraw-Hill, New York, 1969.

2. Taitel, Y., and Dukler, A. E., "A Model for Predicting Flow Regime Transitions in Horizontal and Near Horizontal Gas-Liquid Flow," AIChE J., 22: 47-55 (1976).

3. Hewitt, G. F., "Flow Patterns," in *Two-phase Flow and Heat Transfer,* ed. by D. Butterworth and G. F. Hewitt, Oxford Univ. Press, Oxford, 1977, pp. 18-39.

4. Hewitt, G. F., and Jayanti, S., "To Churn or Not to Churn," Int. J. Multiphase Flow, 19(3): 527-529 (1993).

5. Taitel, Y.; Bornea, D., and Dukler, A. E., "Modelling Flow Pattern Transitions for Steady Upward Gas-Liquid Flow in Vertical Tubes," AIChE J., 26(3): 345-354 (1980).

6. Spedding, P. L., and Nguyen, V. T., "Regime Maps for Air Water Two Phase Flow," Chem. Eng. Sci., 35: 779-793 (1980).

7. Zuber, H., and Hench, J., Rept. No. 62GL1000, General Electric Company, Schenectady, NY, 1962, as cited by Wallis, G. B., *One-dimensional Two-phase Flow*, McGraw-Hill, Inc., New York, NY, 1969, p. 261.

8. Sada, E., et al., "Gas Holdup and Mass-Transfer Characteristics in a Three-Phase Bubble Column," Ind. Eng. Chem., Process Des. Dev., 25: 472-476 (1986).

9. Saxena, S. C.; Patel, D.; Smith, D. N.; and Reuther, J. A., "An Assessment of Experimental Techniques for the Measurement of Bubble Size in a Bubble Slurry Reactor as Applied to Indirect Coal Liquefaction," Chem. Eng. Comm., 63: 87-127 (1988).

10. Zavaglia, J. C., and Lindsay, J. D., "Flash X-ray Visualization of Multiphase Flow During Impulse Drying," Tappi J., 72(9): 79-85 (1989).

11. Kara, S.; Kelkar, B. G., Shah, Y. T.; and Carr, N. L., "Hydrodynamics and Axial Mixing in a Three-Phase Bubble Column," Ind. Eng. Chem. Process. Des. Dev., 21: 584 (1982).

12. Açikgöz, M.; França, F., and Lahey, R. T., Jr., "An Experimental Study of Three-Phase Flow Regimes," Int. J. Multiphase Flow, 18(3): 327-336 (1992).

13. Walmsley, M. R. W., "Air Bubble Motion in Wood Pulp Fibre Suspension," APPITA '92 Proceedings, Australia, 1992, p. 509.

14. Pelton, R., and Piette, R., "Air Bubble Holdup in Quiescent Wood Pulp Suspensions," the Can. J. Chem. Eng., 70: 660-663 (1992).

15. Pan, R.; Bousfield, D.; and Thompson, E. V., "Modeling Particle-bubble Dynamics and Adhesion in Air Bubble/Solid Particle/liquid Systems," Tappi 1992 Pulping Conference, pp. 941-956 (1992).

16. Ellis, J. E., and Jones, E. L., Two Phase Flow Symposium, Exeter, England, June 1965, as cited by Mashelkar, R. A., "Bubble Columns," British Chem. Eng., 15(10): 1297-1304 (1970).

17. Seifert, P., "Recent Innovations in Paper Recycling," Tappi J., 77(2): 149-152 (1994).